TECHNICAL AND ECONOMIC ASSESSMENT OF THE APPLICATION OF MEMBRANE TECHNOLOGY FOR TREATING OIL/WATER/MINERAL EMULSIONS

FINAL REPORT

VOLUME I - LITERATURE REVIEW

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1.0 INTRODUCTION

The total Canadian reserves of heavy oil and oil sands were estimated in 1979 to be 469 billion m^3 (Meyer and Dietzman, 1979), well in excess of the conventional oil reserves. In the coming years production of conventional oil is expected to remain static or decline, while production of bitumen from the non-conventional reserves is expected to increase.

The processes most commonly utilized to recover this resource are thermal processes involving steam injection, which require large volumes of water for steam generation. Depending on the production techniques utilized each barrel of bitumen brought to the surface may be associated with 2 to 10 barrels of water. Since recovery operations in Western Canada are located in water-short or environmentally sensitive areas, effort has been directed in recent years to the development of efficient and economic processes for the recycling of the produced water for treatment and reuse as boiler feedwater.

Several recovery operations are currently recycling the produced water using various combinations of conventional treatment units. To meet boiler feedwater requirements the removal of oil is required, along with hardness and suspended solids removal and silica reduction. One unique characteristic of the produced water from heavy oil/bitumen recovery operations is the frequent occurrence of highly stable oil/water emulsions. The treatment requirements for such emulsions are unique to oil field operations, and concerns have been expressed regarding the reliability, efficiency, and economy of current systems used to break these oil/water emulsions and produce acceptable boiler feedwaters. The need for the development of the efficient and cost-effective recycling technology has been identified by the oil industry. Various alternative treatment and steam generation systems have been investigated to meet this objective. One potential solution to current problems is the application of ultrafiltration and/or reverse osmosis membrane processes for oil recovery and boiler feedwater preparation.

This study was undertaken to provide a technical and economic assessment of membrane technology for treatment of the stable oil/water/mineral emulsions produced during in-situ recovery operations.

Volume I of this report presents the results of a literature review undertaken to review and evaluate the technical efficiency and economics of conventional treatment systems currently employed and assess the potential for membrane applications to meet treatment requirements. Volume II contains the results of an experimental program conducted in order to determine the technical and economic feasibility of the application of membrane technology to emulsion treatment. Volume III contains the Appendices to the report.

To further the literature review appropriate public and private research centers active in oil recovery and membrane research were contacted, and the commercial literature of manufacturers and vendors was examined. Current treatment technologies are discussed, and related industrial applications of membrane systems are assessed. The potential of utilizing ultrafiltration for oil recovery, and reverse osmosis to meet remaining feedwater objectives is discussed.

2.0 **OVERVIEW OF PRODUCED WATER TREATMENT PRACTICES**

2.1 **Produced Water Characteristics**

In recent years, thermal or steam stimulation techniques have been implemented in the recovery of heavy oil/bitumen that is too viscous to flow at normal reservoir temperatures. These in-situ recovery schemes involve heating the oil to lower viscosity and initiate flow. The most common process is the so-called "huff and puff" method which involves intermittent steam injection. Steam flooding, which uses continuous steam injection has been employed as well. Other techniques under investigation include the "combination of forward combustion and water flood" (COFCAW) methods, and combinations of steam and fireflood methods. The formation water and condensed steam which is produced along with the oil, after initial phase separation, is referred to as produced water. Often this water occurs as part of a stable oil/water/mineral emulsion from in-situ recovery operations.

For the purposes of this report "produced water", will refer to the water remaining after separation of free oil. Produced water typically contains emulsified oil, suspended solids, dissolved organics and various inorganic components.

2.1.1 Produced Water Quality

Variations in produced water quality and quantity have been observed between wells and project, as well as over the life of an operation. In steam stimulation, which is a cyclic process, water quality is typically best in the early stages of the production cycle, and deteriorates torward the latter stages. Similarly, poorer water quality in terms of TDS, silica, and hardness is typically observed in subsequent production cycles on the same well (CH_2M Hill, 1982).

The temperature of produced water, which is a consideration in the treatment process, is typically in the range of 65-82°C (150-180°F).

2.1.1.1 Oil/Water/Mineral Emulsions

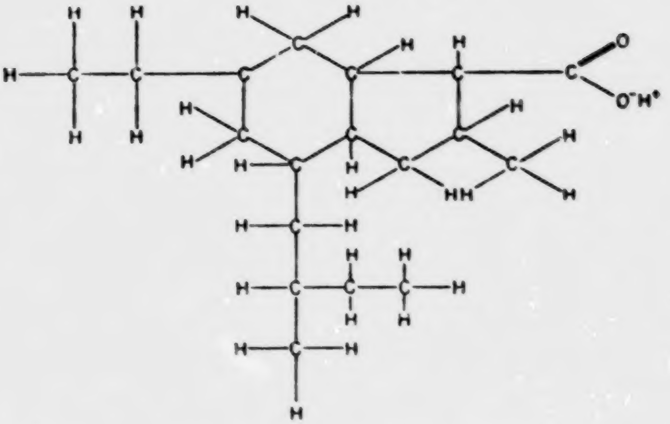
The oil/water/mineral emulsions associated with the production of heavy oil or bitumen are characterized by their unusual stability.

Gewers (1968) investigated the composition of stable heavy oil/bitumen emulsions and found that clay colloids and asphaltenes formed membranous skins which created highly stable emulsions. Conventional de-emulsifying processes, such as use of surface active compounds or electrostatic coalescence were ineffective on clay stabilized emulsions. Gewers found these emulsions to have minimum stability in the pH range of 7 to 9. He also found that the emulsions stabilized by asphaltanic skins are treatable with relatively high concentrations of water soluble surface - active compounds. Typical Alberta heavy oils contain approximately 20 per cent asphaltenes, compared with 2 per cent in conventional light crude. Petroleum acids, such as carboxylic and phenolic acids, can react with caustic soda which is injected to promote recovery, and can result in the formation of surface active chemicals which also promote emulsion stabilization. Figure 2-1 presents the molecular structure of carboxylic and (naphthenic) phenolic acids. Since these are polar compounds, their solubility in water may be a function of pH.

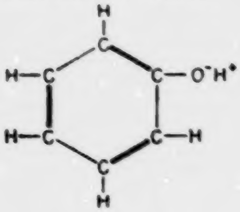
FIGURE 2-1

STRUCTURE OF PETROLIUM ACID EMULSION STABILIZERS
(from Gewer, 1968)

CARBOXYLIC ACID
(napthenic acid)



PHENOLIC ACID



Emulsified oil concentration varies as process conditions change and also varies between fields. At the Cold Lake project for example the emulsified oil in the water after conventional gravity separation normally ranges from about 50 to 400 mg/L. During conditions of treater upset, this level may go as high as 16,000 mg/L (Whalley and Wilson, 1979). Typical suspended oil concentrations in produced waters are reported to range from 100 to 2000 mg/L (CH_2M Hill, 1982).

2.1.1.2 Inorganic Constituents

The inorganic composition of produced water varies significantly between each oil recovery project. Table 2-1 summarizes the major components of produced water from several different pilot studies, and serves to illustrate this variability. A recent survey of heavy oil/oil sand operations in Western Canada (CH_2M Hill, 1982) reported the following range of inorganic constituents:

Total hardness	140-720 mg/L as $CaCO_3$
Silica	100-250 mg/L
Total dissolved solids	6,000-15,000 mg/L
Suspended solids	70-1,000 mg/L
Iron	0.1-0.8 mg/L

2.1.1.3 Organic Constituents

A detailed assessment of organic composition, especially dissolved organics, is not customary for once-through boiler feedwater preparation, and little information is available regarding effects of organics on oil recovery or boiler feed-water preparation. It is

TABLE 2 - 1

A SURVEY OF MAJOR COMPONENTS IN PRODUCED WATER FROM STEAM STIMULATION PROJECTS
(CH_2M - Hill, 1981)

Parameter	Esso Cold Lake Alberta	Suncor Ft. Kent Alberta	BP Marguerite Lake Alberta	Shell Peace River Alberta		Getty Oil Kern. River California	Texaco San Ardo California	Mobil S. Belridge California	Oil McKittrick California	Cat Canyon Field California
				Cold Prod.	Hot Prod.					
Ca (mg/L)	48	290	100	48		-	-	188	109	-
Mg (mg/L)	5	40	30	143	156	-	-	27	56	-
Na (mg/L)	1743	4 500	2500	11 007	2 252	215	2105	3 300	2500	1600
Total Hardness (mg/L $CaCO_3$)	141	890	374	709	-	132	162	580	503	41
CO_3 (mg/L)	15	0	0	300	0	15	0	-	45	0
HCO_3 (mg/L)	304	410	275	6 468	1 558	292	255	-	883	1281
SO_4 (mg/L)	111	525	175	22	379	53	410	-	180	150
Cl (mg/L)	3094	7 253	3850	13 348	2 569	199	2980	4 800	3517	788
S (mg/L)	40	0.01	-	-	-	0	10	-	0	-
NO_3 (mg/L)	-	0.34	-	-	-	0	-	-	-	-
B (mg/L)	10	-	-	-	-	1	-	-	15	-
SiO_2 (mg/L)	250	236	-	8	98	125	130	380	121	178
Fe (mg/L)	0.8	0.1	-	0.3	< 0.1	0.1	1.0	-	-	0.3
TDS (mg/L)	6015	13 300	7300	31 349	7 009	844	6115	13 000	7 000	3 132
pH	8.0	8.1	7.0	8.6	7.1	8.3	6.8	7.2	-	8.2
Conductivity (µmhos/cm)	-	17 000	-	50 000	11 100	1190	-	17 000	8360	4317
Suspended Solids (mg/L)	68	447	-	-		-	69	-	-	130
Oil (mg/L) (Insoluble or suspended oil)	400-2000	220	-	-		200	96	140	-	-

known that the reaction of steam and hot water with bitumen can lead to the formation of water soluble polar organics, and that process chemicals or humic substances may be present in the produced water (Burchfield and Hepler, 1979). Any dissolved substances, including dissolved organics, may be of importance in the recycling or treatment of such water for steam generation under the high pressure conditions commonly utilized. While Table 2-1 does not specify the dissolved organic constituents in the produced waters, it is interesting to note that several publications (Whalley and Wilson, 1979; Imperial Oil, 1978) have given the TOC as >30 mg/L in Cold Lake produced water before treatment. In the case of fireflooding or combustion waters, the TOC before treatment can be near 10,000 mg/L (Barbour and Guffey, 1979).

Produced water from steam stimulation studies at Esso's Leming Cold Lake Area pilot project was analyzed by Tsui and McCort (1979). The analyses of water samples obtained by these researchers were similar to the values shown for Cold Lake in Table 2-1. In addition, they determined NH_3 and phenol concentrations to be in the ranges of 10 to 22 mg/L and 11 to 38 mg/L, respectively. Tsui and McCort also found that, in general, produced water was more toxic to fish than petroleum refinery wastewater or mine depressurization groundwater from the Fort McMurray tar sands area.

A more recent study utilized TOC measurements and GC/MS analysis to identify the organics present in the produced water from two recovery operations in Alberta (ZENON, 1983). The TOC in one sample was appreciable at 583 mg/L. Characterization of the individual trace components in this sample found that in general lower molecular weight substances predominate (<200 a.m.u.), particularly phenol and its alkylated congeners which where present at ppm levels. Also present in lower quantities were benzathiazole, naphthalene and dimethylibenzaldehydes. Variability between sites was found, as the organic content at the second site indicated similar phenol levels, although alkylated phenols and dimethylbenzaldehydes were much lower, while benzathiozole was not detected. Alkylated benzenes were found only in the second site. The study also investigated the effectiveness of various treatment techniques for TOC and trace organic removal. The TOC retentions on DP-1 resin, carbon, and XAD-2 resin were found to be negligible, 40%, and 18%, respectively.

2.1.2 Produced Water Quantity

Thermal recovery operations require large volumes of fresh water for steam generation. Table 2-2 illustrates the water balance at Cold Lake as an example of water requirements at in-situ operation. Total water requirements are in the order of 143,000 m^3/d, of which 35% is to be obtained by recycle.

The ratios of water to oil produced at the Cold Lake Area of Alberta are typically in the area of 2:1. Ratios as high as 10:1 have been reported from some areas of California (CH_2M Hill, 1982). In Canada, water-to-oil ratios are expected to increase with time and forecasters estimate that average water-to-oil and steam-to-oil ratios will increase from about 4:1 in 1984 to 6:1 in the year 2000. The trend, along with projected water quantities produced by year, is illustrated in Table 2-3. Water-to-oil ratios in conventional heavy oil recovery processes also vary between fields, and have been

TABLE 2 - 2

WATER BALANCE AT COLD LAKE (Imperial Oil, 1978)

Water Requirements	10^3 m^3/d
Injection Steam	63.6
Process Water and Steam	19.7
Cooling Tower	25.8
Water Treatment Backwash	17.6
Utility Water	12.7
Potable Water	3.8
Total	143.2
Consumed in Operations	
Displacing Bitumen	25.4
Cooling Tower Evaporation	20.0
Hydrogen Synthesis	2.4
Coke Gasification	5.1
Subtotal	52.9
Waste Disposal	
Deep Wells -	
Water Treatment Backwash	7.6
Sour Water	2.4
Desalter Brine	2.4
Chemical Wastes	0.8
Subtotal	13.2
Surface -	
Water Treatment Backwash	10.0
Cooling Tower Blowdown	3.3
Boiler Blowdown	2.7
Utility Water Waste	10.7
Subtotal	26.7
Recycled	
Produced Water	38.2
Sour Water	4.4
Sanitary Wastes	3.8
Oily Wastes	4.0
Subtotal	50.4
Total	143.2
Water Supply	
Recycled Water	50.4
Fresh Water Make-up	92.8
Total	143.2

TABLE 2 - 3

(ZENON, 1983)

WATER PRODUCTION FORECAST FOR FUTURE STEAMFLOOD AND FIREFLOOD OPERATIONS (Cold Lake, Peace River and Wabasca)

	Steam Huff & Puff*		Steamflood**		Fireflood**	
Year	Steam-to-Oil Ratio	Water (10^3 m^3/d)	Water-to Oil Ratio	Water (10^3 m^3/d)	Water-to Oil Ratio	Water (10^3 m^3/d)
1981	4	8.0	-	-	-	-
1982	4	12.0	-	-	-	-
1983	4	16.0	-	-	-	-
1984	4	16.0	4	4	-	-
1985	4	16.0	4	16	4	16
1986	5	20.0	4	16	3	12
1987	5	20.0	4	16	2	12
1988	5	20.0	4	30	2	15
1989	5	20.0	4	30	2	15
1990	5	20.0	5	37.5	2	15
1991	6	24.0	5	37.5	2	15
1992	6	24.0	5	37.5	2	15
1993	6	24.0	5	37.5	3	22.5
1994	6	24.0	5	37.5	3	22.5
1995	6	24.0	5	37.5	3	22.5
1996	6	24.0	6	45.0	3	22.5
1997	6	24.0	6	45.0	3	22.5
1998	6	24.0	6	45.0	3	22.5
1999	6	24.0	6	45.0	4	30.0
2000	6	24.0	6	45.0	4	30.0

* Steam huff & puff contributes a maximum of 4 x 10^3 m^3/d oil production.

** Steam-to-oil and water-to-oil ratios and trends based on arbitrary estimate and experience.

reported to range from a low of 0.3 to a high of 9.0 (ZENON, 1983). The expected increase in water production with time illustrates the attractiveness of efficient and economical recycling processes, which would eliminate problems associated with obtaining a source of water for steam generation, as well as disposal of unused produced water. Plants recycling water may only partially recycle the produced water, partially recycle and combine the water with other waste streams, or may recycle all the water and supplement with additional water. Thus quantities of produced water to be treated depend on the treatment requirements and process selected at a particular plant.

2.2 Treatment Requirements

The treatment requirements depend on the operating pressure and type of steam generation system used. Operating pressures vary between fields, and reported values range from 700 to 2400 psi (4830-16560 kPa). In the Cold Lake area of Alberta injection pressures typically range from 1200-2000 psi (8280 to 13800 kPa). Generally, higher pressure boilers demand higher quality feedwater. In all water reuse schemes, the type of boiler chosen ultimately determines the cost and complexity of the water treatment required. At the present time, most in-situ pilot or commercial plants use the conventional once-through oilfield steam generators which produce 80% quality steam and require relatively low quality feedwater. A recent survey of steam generation techniques which may utilize produced water found that economics at the present time still favoured the use of these generators (CH_2M Hill, 1982). Treatment requirements are dictated by the need to maximize oil recovery and achieve acceptable boiler feedwater quality for any water being recycled (CH_2M Hill, 1981).

Table 2-4 presents general oilfield requirements for boiler feed water and other uses as obtained from the literature.

To treat produced water for use as once-through boiler feedwater, essentially all oil, hardness and suspended solids must be removed, and a reduction in silica is often required. The removal requirement for Total Dissolved Solids (TDS) varies depending on initial produced water concentration, since levels between fields may vary from less than 1,000 mg/L to over 30,000 mg/L.

In those water treatment schemes in which the produced water is only partially recycled, the remainder is often used for either backwashing filters and softeners or other miscellaneous uses (Stephenson et. al., 1983). Some waste streams which are unsuitable for reuse may be destined for surface or deep well disposal, but environmental regulations still require some degree of treatment for this (Table 2-4). In the event of boiler shutdown, it may be necessary to provide alternate water disposal means, thus unit operations should ideally be able to meet surface disposal guidelines as well.

2.2.1 Oil Removal

The American Boiler Manufacturers Association guidelines recommend oil removal to levels of 7 mg/L or less, with a desirable level being less than 1 mg/L. Oil in concentrations above this can cause scale adherence, film boiling, and coking in the boiler, and

TABLE 2 - 4

WATER QUALITY REQUIREMENTS FOR BOILER FEEDWATER PREPARATION AND OTHER USES

(from CH_2M Hill, 1981)

Parameter	Once-Through Steam Generator Note 4	Thermosludge Generator Note 4	Deep Well Disposal-Note 1	Irrigation	Cooling Water Make-up
Total Hardness mg/L as $CaCO_3$	< 0.5	100-2000			NOTE 2
Silica mg/L	< 50	10-250			NOTE 2
Oil mg/L	< 1	50-1000	< 20	< 5	< 1
TDS mg/L	< 8000	1000-50000		< 1000	-
Suspended Solids mg/L	0	10-500	< 5		NOTE 3
Dissolved O_2 mg/L	< 0.04	-			-
Iron	< 0.05				

NOTE 1 - Water should be chemically compatible with formation waters.

NOTE 2 - The product of Mg as $CaCO_3 \times SiO_2 < 35000$ mg/L.

The product of Ca as $CaCO_3 \times SO_4 < 500,000$ mg/L.

NOTE 3 - Total suspended solids in the recirculating water 200 mg/L.

NOTE 4 - These are vendor recommendations. Some of these values have been exceeded in actual practice.

fouling of ion exchange (softener) resins in the feedwater treatment train.

2.2.2 Boiler Feedwater Requirements

The most common causes of boiler scale are the presence of calcium and magnesium ions. With increasing temperature, the salts of these ions show a decreasing solubility. As a result, most guidelines recommend the removal of hardness to levels of less than 1 mg/L.

Control of silica in process waters is important, since amorphous collidal silica can form glass-like deposits on heat exchange surfaces. In the boiler silica can also produce scaling as a constituent of various complex mineral scales. Silica is not as great a problem in oilfield steam generators as in utility-type boilers, and control may consist of maintaining silica solubility by ensuring that alkalinity is at least 3 times the silica content (but less than 2,000 mg/L), or by careful control of residual calcium levels in feedwater. Operations have been reported with silica levels as high as 150 mg/L (Carrell, 1979).

Oil-field steam generators can tolerate a relatively high level of dissolved solids; their level becomes a concern only when liquid phase concentrations approach solubility limits. Therefore in 80% quality steam generation, feedwater salt concentrations close to 20% of their solubility limits should be possible. Thus sodium chloride levels of up to 50,000 mg/L may be tolerated. Permissible levels of feedwater TDS are generally dependent upon the limitations of the ion exchange softening systems commonly used. Most softeners utilize salt regenerated, strong acid resins whose hardness leakage characteristics limit feedwater concentrations to between 5,000 and 7,000 mg/L. Weak acid resins have been recently applied, and have been reported to operate successfully at 28,000 mg/L, however costs are much higher. The maximum TDS concentration reported in oil-field steam generator blowdown, or condensate film fraction, are 125,000 to 150,000 mg/L (CH_2M Hill, 1982). Practical limits of TDS have been estimated as 40,000 mg/L in the feedwater, and 200,000 mg/L in condensate film.

Suspended solids contribute to softener fouling and boiler sludge formation, and as well they may contribute components in the formation of complex mineral scales. It is therefore critical that boiler feedwater should be essentially free of suspended solids.

Iron should be removed from boiler feedwater due to the possibility of the formation of corrosion products within the generator. Treatment for control of sulfide, oxygen, and pH are also essential to minimize corrosion and to allow successful boiler and equipment operation.

The pH of boiler feedwater must be maintained within the alkaline range, with optimum given as between 8 and 9, to a maximum of 12. These levels will promote silica solubility while also minimizing corrosion. Higher pressure operations generally require a pH of around 10.5 for maximum silica solubility. Values of 13 or higher, however, may result in caustic corrosion or embrittlement, due to excess hydroxide alkalinity.

One proposed system for generation of steam in the oil field is the thermosludge process which was designed to accept produced water directly without the extensive pretreatment required by conventional once-through steam generators. Table 2-4 presented treatment requirements for these generators. The main requirement would be removal of free and emulsified oil, most likely by a combination of skim tank, induced gas flotation and demulsifier addition. This process has been used in the U.S. and Canada, but has not gained commercial acceptance. Water-side scaling, foaming, and oil fouling problems have been reported, and capital and operating costs normally exceed conventional systems (CH_2M Hill, 1982). As well, problems with corrosion due to poor oxygen control have been reported.

Although conventional utility type steam generators are not currently used in the oil-fields, they have several potential advantages over those boilers currently used. The technology for these sytems is highly reliable, they are capable of producing higher quality steam (thus improving recycle efficiency), and they permit the use of solid fuel. The main disadvantage discouraging their application to date is the high cost of water treatment necessary to meet feedwater requirements. Generally, to operate at the pressures required for oilfield operations, these boilers require demineralized feed-water. Table 2-5 illustrates conventional industrial boiler feedwater requirements with variations in pressure. For these boilers, feedwaters must be essentially free of organic compounds as well.

TABLE 2 - 5

(from Simon, 1975)

GUIDELINES FOR WATER QUALITY IN MODERN INDUSTRIAL WATER TUBE BOILERS FOR RELIABLE CONTINUOUS OPERATION[1]

	Boiler Feedwater			Boiler Water		
Drum Pressure (psig)	Iron (ppm Fe)	Copper (ppm Cu)	Total Hardness (ppm $CaCO_3$)	Silica (ppm SiO_2)	Total Alkalinity** (ppm $CaCO_3$)	Specific Conductance (micromhos/cm)
0-300	0.100	0.050	0.300	150	700*	7000
301-450	0.050	0.025	0.300	90	600*	6000
451-600	0.030	0.020	0.200	40	500*	5000
601-750	0.025	0.020	0.200	30	400*	4000
751-900	0.020	0.015	0.100	20	300*	3000
901-1000	0.020	0.015	0.050	8	200*	2000
1001-1500	0.010	0.010	0.0	2	0***	150
1501-2000	0.010	0.010	0.0	1	0***	100

*Alkalinity not to exceed 10% of specific conductance.

**Minimum level of OH alkalinity in boilers below 1000 psi must be individually specified with regard to silica solubility and other components of internal treatment.

***Zero in these cases refers to free sodium or potassium hydroxide alkalinity. Some small variable amount of total alkalinity will be present and measurable with the assumed congruent control or volatile treatment employed at these high pressure ranges.

2.3 Treatment of Oil/Water/Mineral Emulsions

2.3.1 Existing Technologies

2.3.1.1 Design

Oil can be classified by its tendency to separate from the bulk water phase. Free oil is considered to be those large droplets which are readily separable by gravity, dispersed oil describes smaller droplets which may separate due to coalescence, emulsified oil describes oil droplets which do not coalesce or separate, while dissolved oil is that which is soluble in water (Churchill, 1980). Functionally the first three classifications may be distinguished by size. Free oil particles are those larger than 150 microns in diameter, dispersed particles are from 50 to 150 microns in diameter, and emulsions are made up of droplets less than 50 microns in diameter.

A distinction should also be made between oil/water and water/oil emulsions, terms which are often interchanged. Water/oil emulsions, where oil is the continuous phase, generally are treated by well established conventional means; the major difficulty in their treatment is removing the last traces of water. Oil/water emulsions, especially those produced with heavy oil/bitumen, are often very stable and difficult to treat.

For primary oil recovery the well head emulsion is pumped into a treater or series of treater units. Many different types of treaters and/or chemical additions may be used. The simple unit consists of a heated horizontal baffled tank where the initial three phase gas/oil/water split takes place. Treaters may consist of several stages and may include the addition of demulsifying chemicals or a light hydrocarbon diluent. The choice of treater type used and chemicals added is made based on feed characteristics and the quality of the oil and liquid phase streams produced.

Often the stable bitumen-rich oil phase emulsions (water-in-oil) which may be produced during the initial primary separation, may contain up to 40% water. It is reportedly difficult to economically remove sufficient water to meet water content specifications of 0.5% in the product oil. The use of hydrocarbon diluent reduces problems associated with water-in-oil emulsions.

The produced water which typically still contains up to 5000 mg/L oil is then directed to treatment for gross oil/water separation. The most common technique for gross oil/water separation in produced water after the treaters at the present time is gravity separation, such as a skim tank or corrugated plate interceptor. Demulsifying chemicals are commonly used. In general these are high molecular weight alkylene oxide polymers, high molecular weight polyesters prepared from ethylene oxide, or other molecules containing large hydrophobic or hydrophilic portions (McCoy et al, 1982). Following skim tanks, the most popular separation technique for oil recovery has been induced gas flotation, using natural gas, with de-emulsifying chemical addition. Often this treatment is then followed by a polishing filter, utilizing diatomaceous earth, cartridge, or sand filtration.

The actual treatment schematic depends on produced water quality and treatment objectives. Current treatment systems include those designed for deep well injection and recycling for steam generation. Re-injection may only require oil removal to prevent formation plugging, while the more sophisticated recycling systems incorporate a series of unit operations, and may employ combinations of settlers, chemical additions, filters with different media, and softeners. Although several recycling schematics are used, Figure 2-2 prosents what is considered a typical produced water treatment schematic in Alberta.

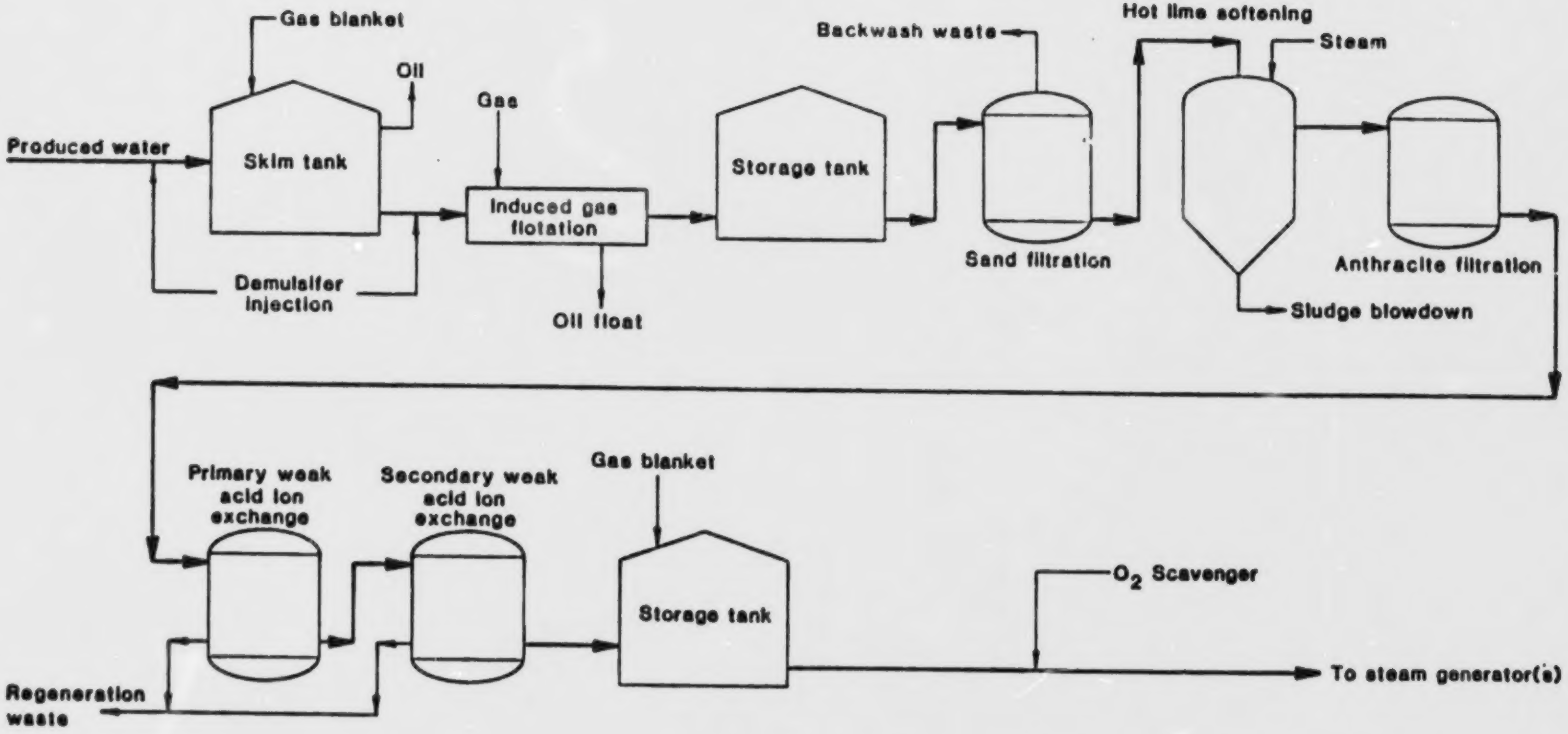
The recovery of emulsified conventional oil from process waters has been researched by many industries, based on payback considerations as well as regulatory requirements which limit effluent concentrations. One such industry is petroleum refining, where environmental regulations strictly limit effluent levels. Typically, free and emulsified oil is treated by dissolved air flotation (DAF) (or similar units) after gravity separation. Luthy et. al. (1978) investigated removal of emulsified oil from refinery wastewaters by the use of various organic coagulants combined with dissolved air flotation. The research found that flocculation with polyelectrolytes prior to DAF may improve operations, but DAF oil removals tended to decrease when suspended solids levels approached 1000 mg/L.

Oil/water separation has also been investigated by the offshore oil industry, and studies have evaluated the use of dissolved air flotation combined with organic coagulants (Snavely and Yarbrough, 1983). Sylvester and Byeseda (1980) studied the process of induced air flotation for oil/water separation in wastewater from all facets of the petroleum industry and found that only oil droplets larger than 2 microns showed significant removal rates by conventional techniques. Dissolved gas flotation has the capability of producing smaller microbubbles than induced gas flotation and hence of producing better

FIGURE 2-2

PROCESS SCHEMATIC FOR TREATMENT OF PRODUCED WATER TO ONCE-THROUGH STEAM GENERATOR FEEDWATER QUALITY

(from CH_2M Hill, 1982)



Gas blanket
Oil
Produced water
Skim tank
Demulsifer injection
Gas
Induced gas flotation
Oil float
Storage tank
Backwash waste
Sand filtration
Hot lime softening
Steam
Sludge blowdown
Anthracite filtration
Primary weak acid ion exchange
Secondary weak acid ion exchange
Gas blanket
Storage tank
O_2 Scavenger
To steam generator(s)
Regeneration waste

oil separation. Application of DGF has been limited in produced waters, due to lack of applications research, higher temperatures which limit gas solubility, and potential equipment corrosion problems. Because of the complexity of these particular emulsions, which are stabilized by colloidal clay, asphaltenes, and resinous acids, conventional emulsion breaking techniques such as those used by other industries may not be effective or economical.

2.3.1.2 Economic Evaluation

The costs of the current technologies for oil removal depend on the process flowrate, and specific units and chemicals required to treat the particular produced water stream, thus an economic assessment is difficult. A recent study by CH_2M Hill (1982), however, developed a cost assessment of what was considered to be a typical produced water treatment process, and their results are presented in Table 2-6. The costs were based on a 73 MW capacity boiler requiring a feed of approximately 500,000 USGPD. The costs developed for the Thermosludge process included removal of oil by a skim tank, with demulsifier addition, followed by induced gas flotation. Thus the capital cost for oil removal, as commonly practiced in the oil field, can be estimated to be in the order of $880,000, in 1982 dollars, or approximately 13% of treatment for a conventional system including oil removal. Operating costs for the oil removal system were estimated at $300,000 per year.

TABLE 2 - 6

CAPITAL AND OPERATING COSTS OF CONVENTIONAL ONCE-THROUGH AND THERMOSLUDGE BOILERS (\$1000)

(from CH$_2$M Hill, 1982)

	CONVENTIONAL ONCE-THROUGH		THERMOSLUDGE	
ITEM	GAS-FIRED	CRUDE-FIRED	GAS-FIRED	CRUDE-FIRED
Capital Cost:				
Boiler feedwater treating facilities	\$ 6 800	\$ 6 800	\$ 880	\$ 880
Steam generator facilities	4 195	4 195	12 100	12 100
Air compression	-	-	-	-
BFW pumps	750	750	-	-
Fuel facilities	-	120	-	120
Miscellaneous	-	80	-	80
Contingency (20%)	2 690	2 730	2 640	2 680
Total	\$14 435	\$14 675	\$15 620	\$15 860
Operating Costs: (Per Year)				
Boiler feedwater treating (including 10% contingency)	\$ 985	\$ 985	\$ 300	\$ 300
Steam generation				
Fuel:				
Natural gas	5 695	-	6 435	-
Crude	-	10 415	-	11 765
Coke/Coal	-	-	-	-
Limestone	-	-	-	-
Maintenance	105	105	435	445
Electrical power	110	115	110	115
Operators	245	245	245	245
Waste disposal and misc.	75	75	125	125
Rental fees	-	-	-	-
Contingency (10%)	620	1 095	735	1 270
Total	\$ 7 835	\$13 030	\$ 8 385	\$14 265

2.3.1.3 Technical Evaluation

Present oil-field steam generation practices use crude oil or natural gas as fuel. Estimates have indicated that on the average, in oil-fired generators using crude oil, one barrel of crude out of every three barrels produced is burned as fuel (Davis, Young and Lyons, 1981). This illustrates that, in addition to other obvious economic considerations, it is critical to recover as much oil/bitumen as is possible. However, due to the stability of the emulsions associated with oil/bitumen production, concern has been expressed regarding the consistent effectiveness of current technologies for breaking these emulsions.

A study by CH_2M Hill (1981) examined potential unit operations for gross oil and fine oil removal which have either been used or considered for the treatment of produced water from heavy oil fields. A summary of their assessment along with other comments obtained from the literature is presented in Table 2-7.

Conventional gravity separation techniques such as the use of skim tanks provide effective gross oil separation. However, examples can be given illustrating some concerns associated with use of these units in heavy oil/bitumen applications.

Konak and Grisard (1980) state that traditional gravity settling techniques (API separation, skim tanks) etc. are inherently slow and inefficient on bitumen and heavy crude produced waters in which oil density is very close to that of water; furthermore high viscosity and surface tension prevent droplet coalescence even after contact has been made. As a result, performance of both gravity settling devices (skim tanks, API separators, plate settlers) and

TABLE 2 - 7

UNIT OPERATION ASSESSMENT OF CURRENT TECHNOLOGIES

(adapted from CH_2M Hill, 1981)

Unit Operation	Advantages	Disadvantages And Constraints
Gross Oil and Solids Removal		
Above Ground Skim Tank	Proven operation, Simple design Low capital cost, Easily covered Provides some surge volume for oil and emulsions. low energy consumption	Limited and variable effeciency Requires large area Sludge withdrawal problems Potentially labor intensive Susceptible to corrosion
Fine Oil Removal		
Induced Gas (Air) Flotation	Proven technology Considerable experience with gas Removes suspended solids as well as oil Can treat emulsion with chemical treatment Less expensive than DAF, Smaller in size than DAF Lower pumping requirements than for DAF	Relatively high energy requirement Chemical handling and storage facilities req'd. Potential float disposal problems Susceptible to process upsets
Dissolved Gas (Air) Flotation	Proven technology with air Can treat emulsion with chemical treatment Removes suspended solids as well as oil	Susceptible to process upsets Relatively high energy requirement Chemical handling and storate facilities req'd. Little experience with gas other than air Potential float disposal problems Efficiency loss at higher temperature
Granular Media Filtration	Well understood technology Wide range of media to suit application Readily available, Simple design Suspended solids removal as well as oil Can be used on high temperatures	Susceptible to process upsets Periodic media replacement Backwashes must be suitabley treated Prone to mud-balling in oily water service
Precoat Filtration	Proven technology, Relatively compact Flexibility in meeting variable conditions Removes solids in submicron range Removes trace quantities of oil Can be completely automated	Spent filter cake disposal required Relatively high capital cost Relatively high operating cost

coalescing devices (coalescers, media filters) would be lower than expected based on experience with paraffinic oils. Brougham (1978) reports the density of bitumen is lower than that of water at temperatures between 37° and 115°C and higher at temperatures below 37°C. While typical oil-field water treatment temperatures after the treaters are above 37°C, a sinking rather than floating oil layer may result if temperatures decrease.

The two most common techniques for fine oil removal, induced gas/air flotation and dissolved air/gas flotation, can treat these emulsions with potentially costly chemical treatment. Generally effluents from these units still contain about 10-20 mg/L oil, with estimates as low as 5 mg/L under optimum conditions (Behie et al, 1982). This level is usually further reduced by some type of subsequent filtration. Some of the problems expressed in the literature associated with filtration are listed in Table 2-7. In addition the oil trapped in the filters is not always recoverable.

There is a lack of operating information available in the literature on the efficiency of oil-field oil and suspended solids removal systems. Thus any evaluation of the technology used must to a large extent be based on an examination of the nature of the wastewater, properties of the oil to be removed, suitability of the unit processes based on experience in other applications, and general knowledge of wastewater treatment technology, design principles, and practices. Complex phenomena are involved in simultaneous removal of emulsified oil and coarse and fine suspended solids with widely varying particle sizes, surface charge, and relative concentrations in the presence of emulsion stabilizing agents. Present practice has apparently evolved from adaptation of existing oil removal and boiler feedwater treatment technology and experience in other applications, initally and mistakenly judged to be similar. This is because, as in many other rapidly developing fields, research and understanding of fundmentals has lagged behind applications.

Induced gas/air flotation and dissolved air/gas flotation are generally suitable processes for oil removal with appropriate chemical addition, and are usually reasonably economic and cost-effective processes. However, several limitations have been identified. Sylvester and Byeseda(1980) found that induced gas flotation was unable to significantly remove oil droplets with a droplet size of 2 microns or less. This implies that a fraction of the emulsified oil will not be effectively removed by these units. Oil droplet and gas bubble diameters, liquid residence time, and cationic polyelectrolyte dosage were the most significant variables affecting overall performance. Of these factors, polyelectrolyte dosage is the only one easily changed in an operating system running at design throughput.

The deficiencies of currently utilized unit processes appear to be of several types. Firstly individual unit operations are inherently unable to achieve the necessary levels of efficiency to permit reducing the number of treatment stages. For oil and suspended solids removal a maximum of two process steps, for gross oil and solids removal, and fine oil and solids removal would appear desirable, but current treatment trains utilize more steps at lower individual efficiences. Secondly, some individual processes are only marginally suitable for the service selected, and are inherently unable to handle surges in inlet concentrations or hydraulic loadings without loss of efficiency. Because of the low rate of separation of heavy oil in gravity devices, hydraulic surges can potentially result in serious losses in efficiency and high outlet oil concentrations.

Experience in other applications has shown that granular media filters, although effective for removal of turbidity, are unable to remove high oil concentrations. In granular media, oil removal is a complex function of direct filtration, coalescence, and adsorption onto filtered solids. While oil removal capacity of media filters is quite limited, it may be enhanced by a variety of techniques, including the use of rougher media such as anthracite and by addition of a coagulant to provide flocculant solids trapped within the filter media onto which oil droplets can adsorb. Nonetheless sudden high concentrations (slugs) of oil tend to break-through filter media into the effluent. This is often followed by mud-balling of the fouled media and associated backwashing problems.

Finally process control technology for individual unit processes is generally relatively primitive, and does not permit automated feed-back or feed-forward control, based on measurement of changes in key process operational parameters, to head off losses in efficiency. This is a general problem in many areas of wastewater treatment, and control is most desirable in those processes in which outlet concentrations respond most strongly to changes in inlet concentration. In coagulation, for example, the state-of-the-art does not permit reliable automated adjustment of coagulant dosage based on resultant tubidity. A similar type of control is desirable but non-existent in induced air flotation units. A recent laboratory study carried out on removal of oil from produced water by gas flotation (Strickland, 1980) concluded that no correlation between basic system properties and attachment efficiency of gas bubbles to oil-droplets could be found. Furthermore while some important parameters of performance were determined by cell design, others were characteristics of the feed. Therefore the same unit will produce different oil concentrations on different feeds, and effluent from a given cell would change if feed characteristics change. Salinity, oil droplet size, and oil concentration all influenced performance. Because fundamentals are so poorly understood it is premature to discuss better process control in these units. Another recent laboratory study (Behie et al, 1982) concluded that while very high efficiencies were potentially achievable with induced gas flotation, the improper design of only one important element of a unit (e.g. air sparger) can result in very poor performance even if all other elements are well designed. Problems can be expected to be particularly acute whenever feedwater concentrations change severely and rapidly.

2.3.2 Emerging Technologies

Numerous alternative unit processes are being investigated and some have been piloted by the oil industry. API separators, common for the treatment of refinery wastewaters, and corrugated plate interceptors have been investigated for gross oil removal. Although both these techniques may show potential for free oil removal, they are not effective on emulsions.

Chemical coagulation is one alternative process which has received very limited study for treatment of emulsified oil. In particular one potentially applicable technique is to utilize coagulant flocs as oil adsorbents then use DAF to separate the oil-laden flocs from the oily water and subsequently recover oil from the flocs. This technique is widely used at large scale in the primary steel and aluminum industries and automotive industry for treatment of cold rolling and stamping emulsions respectively.

One option under investigation is the use of coalescers, based on adsorption of the oil onto a solid surface. Many types of media have been successfully applied in coalescers and include; granular polymeric resins, specially treated sand, and polyurethane foam. In general the application of coalescers for the treatment of produced water has been limited in the past by a number of factors. Traditional coalescers are designed for treatment of oily wastewaters with negligible suspended solids concentrations operated in an upflow configuration; however, high concentrations of suspended solids cannot be handled by this configuration as the solids will plug the media and cannot be removed by backwashing. Furthermore, very stable emulsions caused by relatively large quantities of surfactants often must be pretreated by the addition of de-emulsifiers. Another factor that has limited the application of coalescers is that many of the traditional coalescing media will not tolerate the higher temperatures of produced waters. As a result it seems likely that applications would be limited to use as secondary or polishing filters. Further development of these technologies will depend upon the development of economical oil recovery processes and the application of inexpensive or easily regenerated adsorbents.

Extensive research is being conducted in the development of alternative techniques for breaking of bitumen/water emulsions. In an effort to assess alternate processes of emulsified oil separation, Konak and Grisard (1980) examined deoiling of Cold Lake produced water by acidification. Although chemical demusifiers were in satisfactory operation at Cold Lake, the process was examined as a back-up technology directed at situations where the dissolved organic content was a concern and to assist in reducing hydrogen sulfide concentration. They found that acidification with a variety of acids, including sulfuric, hydrochloric, nitric, and acetic, to pH of between 4 and 4.5 was successful in deoiling produced water, as well as in reducing dissolved organic matter and sulfide more effectively than with demulsifiers.

Nagendran and Hrudey (1981) investigated the use of electrolytic flotation techniques in treatment of emulsified oil in Cold Lake produced water. Although the treatment showed promise, costs of required chemical coagulants were prohibitive.

Demulsifiying chemicals have been studied extensively. A process patented by Farnand (1983) found that addition of a combination of formic acid, oxalic acid, and SO_2 was effective in breaking oil/water emulsions typical of in-situ bitumen/heavy oil operations. A variety of organic and inorganic acids, salts, and commercial demulsifiers where tested, of which formic acid was most effective with least side effects. Some adverse effects of using mineral acids have been claimed to be clay attack and associated flocculation. The reasons for the particular effectiveness of formic acid for this application are not fully understood. McCoy et. al. (1982) patented a process for breaking bituminous emulsions using water soluble demulsifiers consisting of ethylene oxide and propylene oxide polymers (molecular weights greater than about 10,000), at temperatures between 25 and 120°C. Several other patents have been filed in the U.S. using various combinations of polyols, with and without other additives. A Texaco patent (U.S. Patent No. 4,058,453) described a similar process for breaking tar sand emulsions with high molecular weight ethylene oxides and the optional addition of an alkali earth metal.

A review of polar demusifying agents for heavy oil emulsions by Farnand (1981) found aliphatic polar compounds containing oxygen in their functional groups and with a molecular weight of 80 to 200 were most effective. Aliphatic acids, alcohols and ketones were proposed as acceptable additives because they were inexpensive, non-toxic, biodegradable, and readily recoverable.

An investigation of emulsions from the steam injection operation of Norcen Energy Resources, was undertaken by Majid and Farnand (1980). Numerous additives were tested and evaluated, and the authors found that the emulsions were treatable with suitable additives, and water content was reduced from about 70% to below 1%.

One proposed process of breaking oil/water emulsions from in-situ oil recovery involves adding the emulsion to an aqueous coal slurry (Rowat, 1980). The oil attaches to the surface of the coal particle while water is rejected due to the hydrophobic nature of the coal.

A patent for breaking small emulsions of a diameter at least 0.02 microns, was filed by Bitter and Reynhout of Shell International Research (1980). The process investigated a micro-pore ultrafiltration membrane, consisting of hydrophilic and oleophilic zones whose pore size diameter is at least twice the diameter of emulsified droplets, with a recommended maximum pore size of 10 microns. The process functioned not by concentrating the oil in a concentrate stream but by providing surfaces for the coalescence of the emulsified oil to take place. The entire flow is passed through the membrane and is followed by a gravity separation of the two phases.

In recent years, the potential of ultrafiltration membranes for treatment of bitumen/water emulsions has been indicated. Although little published information specific to this application is available, some promising research has been undertaken. Application of ultrafiltration to the treatment of other industrial emulsified oily wastes has been proven to be efficient and economical. A more detailed discussion on this emerging technology will be presented in section 3.0.

2.4 Preparation for Use As Boiler Feedwater

2.4.1 Existing Technologies

2.4.1.1 Design

In addition to removal of oil/bitumen, produced waters must meet all other objectives for boiler feedwater, as discussed in section 2.2.2. Processes for boiler feedwater preparation are used routinely in many industries, although flow schematics may vary from those used in heavy oil recovery operations.

For hardness removal, the most common technique is ion exchange softening. The type of resin chosen depends on the upstream total dissolved solids level. The Esso Resources Cold Lake project is one large - scale system which employs hot lime softening for simultaneous reduction of both hardness and silica.

Getty Oil Company, in the Kern River Field near Bakersfield, California, operates one of the largest produced water recycling systems for production of boiler feedwater and treats approximately 80,000 m^3/day of produced water (Carrell, 1979). The treatment successfully employed consists of chemical treatment, induced gas flotation (depurator unit), sedimentation, nitrogen gas flotation, pressure cake filtration (diatomaceous earth), and cation exchange softening. Figures 2-3 and 2-4 illustrate the water quality treatment sequence, and unit removal efficiencies. The operators claimed that oil concentration was one of the most critical factors in plant operations due to the adverse effect on filter operations. Removal of 90% of the oil is accomplished by the initial IGF (depurator) in a single step while removal of the remaining 10% requires a series of steps. Removal of suspended solids requires a series of steps. Removal of hardness is achieved by the simple step of cation exchange softening. The multiple steps involved for removal of oil and suspended solids indicate an undesirable lack of efficiency in each stage. The use of diatomaceous earth pressure filtration is also generally indicative of the presence of colloidal, sticky, gelatinous or other difficult to remove materials in the water to be

FIGURE 2 - 3

WATER TREATMENT PROCESS AT KERN RIVER
(from Carrell, 1979)

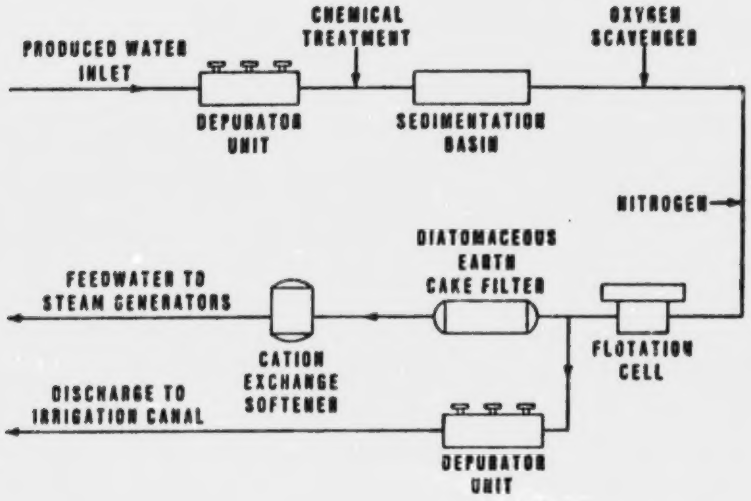
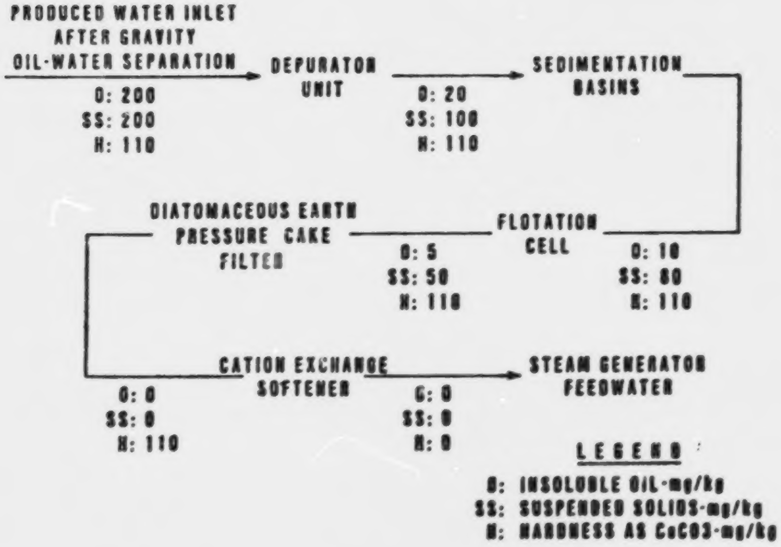

PRODUCED WATER INLET
CHEMICAL TREATMENT
OXYGEN SCAVENGER
DEPURATOR UNIT
SEDIMENTATION BASIN
NITROGEN
DIATOMACEOUS EARTH CAKE FILTER
FEEDWATER TO STEAM GENERATORS
CATION EXCHANGE SOFTENER
FLOTATION CELL
DISCHARGE TO IRRIGATION CANAL
DEPURATOR UNIT

FIGURE 2 - 4

UNIT REMOVAL EFFICIENCIES AT KERN RIVER
(from Carrell, 1979)


PRODUCED WATER INLET AFTER GRAVITY OIL-WATER SEPARATION
O: 200
SS: 200
H: 110
DEPURATOR UNIT
O: 20
SS: 100
H: 110
SEDIMENTATION BASINS
O: 10
SS: 80
H: 110
FLOTATION CELL
O: 5
SS: 50
H: 110
DIATOMACEOUS EARTH PRESSURE CAKE FILTER
O: 0
SS: 0
H: 110
CATION EXCHANGE SOFTENER
O: 0
SS: 0
H: 0
STEAM GENERATOR FEEDWATER
LEGEND
O: INSOLUBLE OIL-mg/kg
SS: SUSPENDED SOLIDS-mg/kg
H: HARDNESS AS CaCO3-mg/kg

treated, and results in loss of the oil removed with the discarded diatomaceous earth.

The process developed for water recycling in Cold Lake is shown schematically in Figure 2-5. The key elements of this process are deoiling (by induced gas flotation), softening (by hot lime softeners), and deionization by ion exchange. As shown in this Figure, make-up water is combined with produced water prior to entering the treatment process.

Whalley and Wilson (1979) discuss the various other treatment process schematics which were evaluated for the Cold Lake project before selection of the system described previously. These are presented in Figure 2-6, with technology assessment shown in Table 2-8. Three potential schemes investigated utilizing drum-type utility boilers. In one scheme (Figure 2-6c) desalination (by evaporation or membrane techniques) of produced water was to be followed by demineralization of combined pre-treated produced water and make-up water. The other two schemes using utility boilers handled the treatment of produced water in a relatively conventional manner, and terminated with a softening step. Introduction of make-up water prior to softening helps in reducing hardness leakage.

A recently expanded thermal pilot project operated by Suncor Inc. near Bonneyville, Alberta is expected to require 3,700 m^3/day of steam generator feedwater (Kloepfer and Kus, 1983). However, the fact that the produced water contained TDS levels up to 20,000 mg/L (versus approximately 6,000 in Cold Lake) resulted in a decision to recycle only up to 50% of the produced water, and supplement it with local municipal wastewater. The processes selected for treatment of this

FIGURE 2 - 5

PROCESS SCHEMATIC FOR RECYCLE OF PRODUCED WATER FROM CYCLIC STEAM STIMULATION BITUMEN RECOVERY AT COLD LAKE (Imperial Oil, 1978)

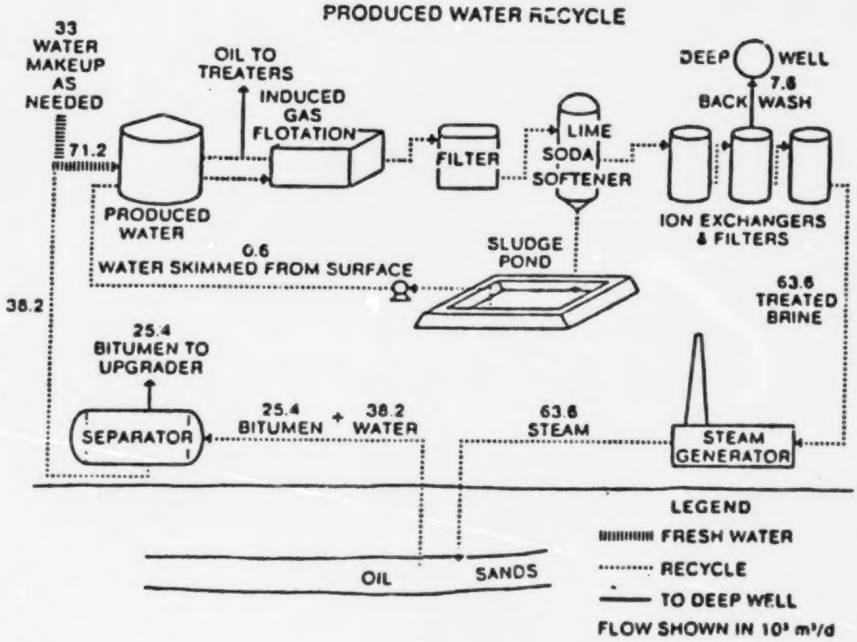
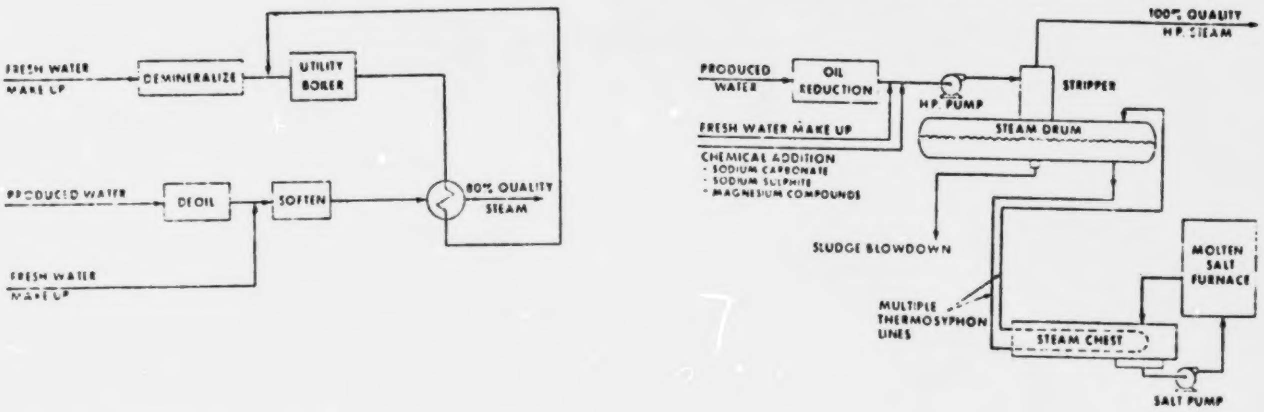

PRODUCED WATER RECYCLE
33
WATER
MAKEUP
AS
NEEDED
71.2
OIL TO
TREATERS
INDUCED
GAS
FLOTATION
FILTER
LIME
SODA
SOFTENER
DEEP
WELL
7.8
BACK
WASH
ION EXCHANGERS
& FILTERS
PRODUCED
WATER
0.6
WATER SKIMMED FROM SURFACE
SLUDGE
POND
38.2
63.6
TREATED
BRINE
25.4
BITUMEN TO
UPGRADER
SEPARATOR
25.4
BITUMEN
+
38.2
WATER
63.6
STEAM
STEAM
GENERATOR
OIL
SANDS
LEGEND
FRESH WATER
RECYCLE
TO DEEP WELL
FLOW SHOWN IN 10³ m³/d

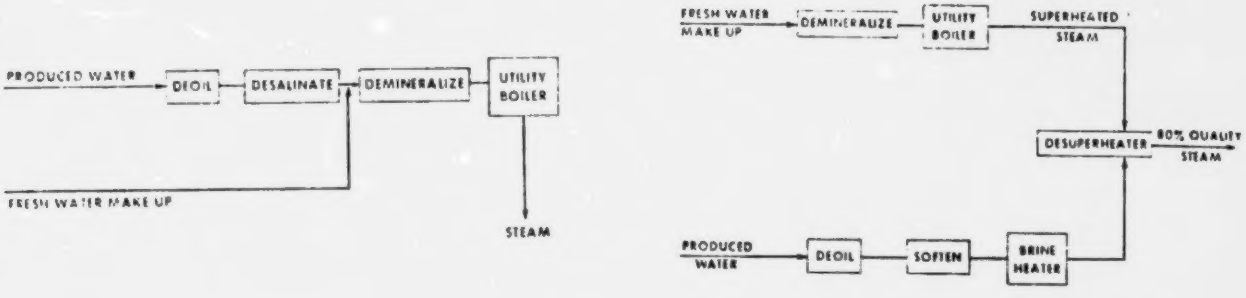
FIGURE 2 - 6

POTENTIAL WATER RE-USE SCHEMES
(from Whalley & Wilson, 1979)


a) REBOILER REUSE SCHEME

b) THERMOSLUDGE GENERATORS

FRESH WATER MAKE UP
DEMINERALIZE
UTILITY BOILER
PRODUCED WATER
DEOIL
SOFTEN
80% QUALITY STEAM
FRESH WATER MAKE UP
100% QUALITY H.P. STEAM
PRODUCED WATER
OIL REDUCTION
H.P. PUMP
STRIPPER
FRESH WATER MAKE UP
STEAM DRUM
CHEMICAL ADDITION
- SODIUM CARBONATE
- SODIUM SULPHITE
- MAGNESIUM COMPOUNDS
SLUDGE BLOWDOWN
MOLTEN SALT FURNACE
MULTIPLE THERMOSYPHON LINES
STEAM CHEST
SALT PUMP


c) UTILITY BOILER REUSE SCHEME

d) SUPERHEAT/DESUPERHEAT REUSE SCHEME

PRODUCED WATER
DEOIL
DESALINATE
DEMINERALIZE
UTILITY BOILER
FRESH WATER MAKE UP
STEAM
FRESH WATER MAKE UP
DEMINERALIZE
UTILITY BOILER
SUPERHEATED STEAM
DESUPERHEATER
80% QUALITY STEAM
PRODUCED WATER
DEOIL
SOFTEN
BRINE HEATER

TABLE 2 - 8

ASSESSMENT OF POTENTIAL WATER REUSE SCHEMES EVALUATED FOR COLD LAKE

(Whalley & Wilson, 1979)

Potential Water Reuse Schemes.

Scheme	Water Treatment	Steam Generation
1.	Deoil, Desalinate, Demineralize	Utility-Like Boilers—100% Make Up
2.	Deoil, Soften	Utility Boilers—Reboilers
3.	Deoil, Soften	Utility Boilers—Superheat/Desuperheat
4.	Oil Reduction	Thermosludge Generators
5.	Deoil, Soften, Silica Reduction	Conventional Steam Generators

Relative Merits of Reuse Schemes.

Attribute	Utility Boilers	Reboilers	Superheat /Desuperheat	Thermosludge	Conventional Steam Gen.
Flexibility	Limited	High	High	High	High
Cost	Very High	High	High*	Moderate	Low
Technical Certainty	Moderate	High	Moderate	High	High
Ability to Fire Coal	High	High	High	Low	Low

*High if coal-fired, only moderate if gas-fired.

stream include chlorine disinfection, lime/soda softening, filtration, and ion exchange. Operating data has not been reported in the open literature.

2.4.1.2 Economic Evaluation

For later comparison with UF/RO treatment, total capital cost from Table 2-6 of a complete treatment system based on a flow of approximately 500,000 USGPD is estimated to be $6,800,000 ($CH_2M$ Hill, 1982). Operating costs were estimated at $985,000. In section 2.3.1.2, the capital and operating costs of typical conventional systems for oil removal were estimated at $880,000 and $330,000 per year, respectively. Substracting these from the cost of total treatment yields a rough estimate of the cost of processes for achieving boiler feedwater quality after oil removal. Therefore the capital and operating costs for a process employing sand filtration, hot lime softening, anthracite filtration, and primary and secondary ion exchange are estimated at approximately $6 million and $700,000 respectively.

2.4.1.3 Technical Evaluation

Although information on the design of typical unit processes can be obtained, there is little information published regarding the effectiveness or efficiency of water recycling systems presently in use in the oil-field. Problems may be encountered in the following treatment stages when used:

i) Lime softening (in particular when there is a high silica content) and lime sludge disposal.

ii) Ion exchange fouling due to organic content.

iii) Boiler scaling

iv) Corrosion of piping and other equipment.

Very high demands are placed upon the technology for recycling in-situ water and, therefore, the likelihood of problems relating to the above areas is also high, especially during the initial start-up phases. Although the Cold Lake project has proven the technical feasability of recycling produced water, concerns have been expressed in the industry regarding cost, reliability, and process control (Card and Kus, 1983).

The main technologies in current use for the recycling of produced water from heavy oil/in-situ bitumen operations have been adapted from classical boiler feedwater treatment principles as used in other industries. The main adjustments made are a consequence of the higher levels of TDS (brine), dissolved gases, and silica encountered. The major strength of existing systems is that not withstanding some problems systems are operational and functional for the recovery of oil and generation of boiler feedwater. Current technologies have permitted substantial amounts of produced water to be treated and recycled for use in once-through steam generators.

The limited information available on their operating efficiencies has indicated that these systems are generally satisfactory with respect to treatment objectives, however many systems are complex and expensive to operate. They often require a large number and variety of specialty chemicals, a high level of process monitoring, skilled service by water treatment chemical suppliers, and special maintenance procedures to keep equipment operational. Additionally large amounts of commodity chemicals are used which may generate significant quantities of sludge or liquid wastes requiring subsequent disposal. These include lime for softening and brine for ion exchange regeneration. A frequently occurring but not widely publicized problem with the type of treatment train commonly used is that water-soluble polar organics are prone to fouling weak acid ion-exchange resins, leading to loss of exchange capacity.

Greater simplicity is desirable in oil-field treatment systems. The limited recycle practised to-date is a result of the economical and technological limitations of current systems, and suggests that more cost-effective processes are needed.

2.4.2 Emerging Technologies

The technologies currently utilized are various combinations of well established conventional processes similar to those utilized by other industries. One technology offering considerable potential for this application is the use of reverse osmosis for boiler feedwater preparation. Section 3.0 discusses membrane processes in general, and related industrial applications of membrane technology which serve to illustrate the potential efficiency and cost-effectiveness of membrane processes for treatment of oil-field produced waters.

3.0 APPLICATION OF MEMBRANE TECHNOLOGIES FOR TREATMENT OF PRODUCED WATER

3.1 Background

In recent years, both research and applications of membrane technologies for the treatment of various waters and industrial wastewaters have increased. With improvements in operating efficiences, widespread commercial acceptance of large scale systems has also occurred.

The commonly used pressure driven membrane process are ultrafiltration (UF) and reverse osmosis (RO). Ultrafiltration is a low pressure membrane process, suitable for separation of particulate and colloidal material and large dissolved organic molecules. Reverse osmosis is a high pressure membrane process utilized primarily for separation of ions and smaller dissolved organic molecules. In both cases, separation is accomplished by selective passage of components through a membrane with specific characteristics. A relatively pure "permeate" passes through the membrane and is collected, whereas rejected materials are concentrated in the "concentrate" or "retentate".

Table 3-1 summarizes and compares the characteristics of these two processes, and illustrates membrane separation ranges and process effectiveness compared to conventional treatment.

The treatment of produced water for use in once through steam generators requires removal of essentially all oil, hardness, and suspended solids, and often reduction in silica and dissolved solids. The use of ultrafiltration for removal of emulsified oil from industrial waste streams has been well documented, and large scale

TABLE 3 - 1

SUITABILITY OF PROCESS

Contaminant	Particulate Matter			Colloids	Dissolved Inorganics (ions)	Dissolved Organics		Membrane Pore Size	Typical Applied Pressure	
Process	>10 u	>2 u	>.2 u	≥100 Å		>10,000 MW	>100 MW	A°	psi	kpa
Ultrafiltration	Excellent	Excellent	Excellent	Excellent		Excellent	Fair	50 - 500	20-100	138-690
Reverse Osmosis		Excellent	Excellent	Excellent	Good	Excellent	Good	5 - 50	200-800	1380-5520
Inert Media Filtration	Excellent	Good								
Adsorption	Excellent	Good				Excellent	Excellent			
Microfiltration	Excellent	Excellent	Good							
Deionization					Excellent					

reverse osmosis units are now commonplace for treatment of seawater, boiler feedwater, and metal finishing waters. The technical and economical advantages of these related applications illustrate the potential for UF/RO application in the treatment of oilfield produced waters for steam generation.

3.2 Treatment of Oil/Water/Mineral Emulsions by Ultrafiltration

3.2.1 Process Description

Ultrafiltration is basically a low pressure membrane filtration process on the molecular level by which solvent and solute molecules are separated by shape and size with some degree of influence from ionic strength and the pH of the solution. Ultrafiltration membranes typically have a pore size of 50 to 500 angstroms.

Figure 3-1 shows the process of passing a feed solution under turbulent conditions, under pressure across the UF membrane surface. In this process the small molecules pass through the pores of the membrane and as the feed solution passes over the membrane surface it becomes more and more concentrated with large molecules.

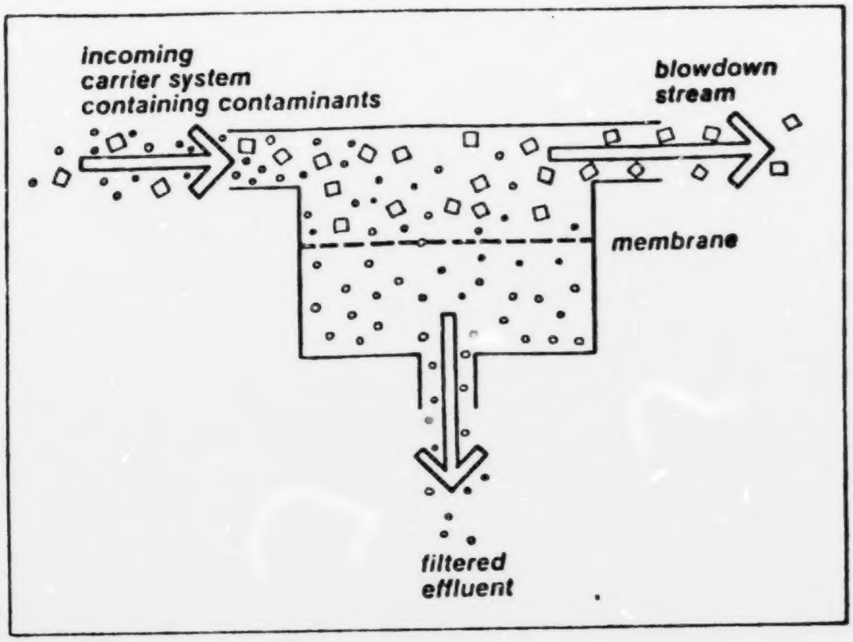
Permeate rate is a function of applied pressure, temperature, turbulent pumping flow, steric effects, and the degree of concentration polarization. Operating pressures typically range from 20-100 psig (138 - 690 kPa). Flux, a term used to describe membrane performance, is defined as the rate of permeate flow per unit area of membrane per unit time.

Ultrafiltration can be used to selectively concentrate valuable components from wastestreams, and has been commercially

FIGURE 3-1

MEMBRANE ULTRAFILTRATION PROCESS

Wastewater is pumped under pressure to the semi-permeable membrane which permits passage of small molecules but retains the large ones.


incoming
carrier system
containing contaminants
blowdown
stream
membrane
filtered
effluent

accepted as a treatment for oil/water emulsions. The open literature contains information on systems upwards of 100,000 gpd in size which have been successfully employed for this application. Depending on membrane type and configuration, systems can be designed to totally remove flocculants or suspended solids, as well as colloids and turbidity.

The stability of oil/water emulsions is enhanced by the prescence of surfactants, which lower the interfacial tension and reduce oil coalescence. Surfactants may be naturally occurring compounds, compounds added to promote emulsion stability at other stages of the process or by-products of industrial processes. The surfactant molecule generally consists of a long chain compound with an oil soluble, non-polar "head" and a large water soluble, polar "tail".

Ultrafiltration separates compounds on the molecular scale, and in industrial oily water applications, a varying proportion of the smaller surfactant molecules may pass through the membrane, while the larger oil particles would not. The loss of the emulsion stabilizing agent with the water in the permeate could therefore cause the oil/water emulsion to break down and release free oil. Depending on the membrane affinity for oil this may result in formation of an oily layer on the membrane surface, accompanied by a gradual decrease in permeate rate with time. However, the rate can often be restored to its original value with cleaning at appropriate intervals. It is also possible that the oil droplets remaining during concentration of the emulsion may coalesce due to the increasing oil concentration and not adhere to the membrane surface. Many applications have achieved up to

90% volume reduction, thereby concentrating the oil by a factor of 10. The concentrate with accumulated free oil can be removed and often phase separation actually occurs. Three layers containing recoverable oil, oil/water, and sludge may be separate.

Emulsions from in-situ steam stimulation operations can be highly stable; this stability is thought to be associated with the presence of colloidal asphaltenes and clays, as well as various petroleum acids. Colloidal compounds do not pass through ultrafiltration membranes, and passage of petroleum acids, such as carboxylic acids which are relatively small in molecular weight, is dependent on their affinity to the colloidal suspensions present. However, previous ultrafiltration research on produced water emulsions (Farnand and Poirier, 1983) has suggested these emulsions break at the membrane surface thus implying destabilization may occur due to surfactant or emusifier loss into the permeate. The study indicated the potential for ultrafiltration to break these emulsions, but found that effectiveness is dependent upon the membrane material used.

3.2.2.1 Membrane Materials

To prevent membrane pore clogging by oil, and to allow water to pass through the membrane, hydrophilic and oleophobic ultrafiltration membranes are usually selected for treatment of oily wastes.

There are various ultrafiltration membrane materials commercially available and under development for treatment of oil/water emulsions. Table 3-2 lists the commercially available ultrafiltration membranes, and includes pH and temperature

TABLE 3-2

COMMERCIAL ULTRAFILTRATION MEMBRANES

Membrane Material	Configuration	MW Cut Off	Permeate Rate ($L/m^2/d$)	Pressure (kPa)	pH Range	Temp. (°C)
Cellulose acetate	Spiral	15000	6000	360	3-7.5	50
	Tubular	1000-100000	400-10000	310	2-8	35-50
	Flat	1000-100000	400-10000	310	208	35-50
Co-Polyacrylonitrile/ Methallyl Sulfonate	Flat	20000	7000	206	1-10	40
Modacryl Polymers	Capillary	50000-80000	12000-15000	103	1.5-13	60
Polyacrylonitrile	Tubular	10000	7000	310	4-10	60
Polybenzimidazolone	Tubular	NA	NA	NA	1-12	60
Polyamide	Capillary	2000-50000	100-6000	155	2-12	80
	Tubular	8000-200000	3000-9000	310	2-8	40
Polyolefin	Spiral/Tubular	20000-100000	NA	NA	2-10	40
Polysulfone	Flat	1000-2000	4000-12000	360	1-13	90
	Spiral	1000-80000	2000-4000	516	2-14	50
	Capillary	2000-50000	1000-14000	103	1.5-13	75
	Tubular	5000-50000	8000	310	1-13	40-95
Polyvinylidene FLuoride	Spiral	10000-20000	7500	360	2-12	90
Sulfonated Polysulfone	Flat	8000-20000	1000-15000	310	2-12	80
ZrO_2	Tubular	20000	NA	NA	1-14	>93
Non-cellulosic Hydrophilic Polymeric	Tubular	10000-20000	8150-16000	710	1-14	40
	Flat	5000	5000	890	1-14	100

(adapted from Purch and Welch, 1982)

limitations, and approximate permeate rates based on pure water fluxes for each membrane type.

There are basically three types of commercially available ultrafiltration membranes; cellulose acetate derivatives, polysulphones, and a limited class of other non-cellulosic polymeric membranes. In addition to the commercially available membranes, researchers are continually improving upon and developing new membrane materials. However data is generally not available on many recently developed membranes due to their proprietory nature.

The literature has shown that many companies have demonstrated the efficient and economic concentration and recovery of oil (mineral and vegetable) from oil/water emulsions utilizing ultrafiltration systems. ZENON Environmental Inc. and others, have industrial systems in the automotive industry treating drawing, stamping, and machinery oil emulsions for one or more of the following purposes:

i) To reduce emulsion volume,
ii) To reclaim oil for reuse, or
iii) To produce a permeate suitable for disposal or reuse.

Typically, these systems can concentrate oily waste streams to 80-90% volume reduction up to a 30-40% oil content. Ultrafiltration is also very effective for removal of suspended solids.

To meet one or more of the objectives listed above, the membrane should preferably be hydrophilic. Commercially available hydrophilic membranes include cellulose acetate (CA), polyvinylene difluoride (PVDF), sulphonated polysulphone (SPS), and thin-film composite (TFC) formulations of these types of membranes. Ultrafiltration membranes such as polysulphone and polyimide, due to their oleophilic nature, can be fouled by oil and hence are usually not thought to be suitable for processing of oil-in-water emulsions. However, the chemical properties of the feed water in combination with chemical composition of the specific membrane may result in surface interactions which cause a decrease the expected fouling rate. Thus laboratory tests are necessary to determine which actual membrane materials are suitable for a particular application.

Reverse Osmosis membranes due to their smaller pore size and pressures used are susceptable to fouling by oil laden feedwaters. No published literature could be found for treatment of oily waste water by reverse osmosis. In addition, all membrane suppliers contacted except for one indicated that all oil must be removed before employing their RO membranes. One manufacture has indicated that a newly developed sulphonated polysulphone reverse osmosis, hollow fibre membrane might be capable of concentrating oil/water solutions.

In addition to membrane characteristics, operating conditions will affect membrane selection. Ultrafiltration membranes are affected by a number of factors which limit the conditions under which they can be operated. These factors include:

i) pH
ii) temperature
iii) resistance to bacteria
iv) resistance to chlorine
v) resistance to solvents

Table 3-3 presents these limitations for five membranes that are viable for oil/water separation.

Generally the pH and temperature limitations for SPS, PVDF, and TFC membranes are due to the deterioration of the membrane housing material. Cellulose acetate membranes may also be degraded by bacteria, and dissolved by organic solvents. Cellulose (CE) membranes are a new class not yet commerically available, however, they offer a greater working pH range and improved organic rejection than cellulose acetate.

Produced water is generally brought to the surface at temperatures above 65°C (150°F). Depending on actual temperatures in the field treatment system, this may require cooling if the cellulose acetate or TFC membranes are utilized. In the case of other membranes, care must be taken to exclude all heat sensitive materials from the membrane module. Current commercial membrane modules include such materials, and therefore, as shown in Table 3-3, emulsion fed to these modules would also require some cooling. The pH of the water is generally slighly alkaline, and well within the acceptable range of the ultrafiltration membranes listed.

3.2.1.2 Configurations

In addition to membrane type, there are several configurations of membranes available. Configurations are selected according to availability, feedwater quality, pretreatment requirements associated with each configuration, and economical considerations. The most common membrane configurations are tubular,

TABLE 3-3

TYPICAL OPERATING LIMITATIONS OF SOME HYDROPHILIC ULTRAFILTRATION MEMBRANES

	CA	PVDF	SPS	TFC	CE	PS	CHP
pH	4-7	2-12	2-12	2-10	4-12	2-12	2-12
Maximum Temp. (°C)	35	60	60	35	35	60	60
(°F)	95	140	140	95	95	140	140
Bacterial Resistance	no	yes	yes	yes	no	yes	yes
Chlorine Resistance	yes	yes	yes	no	yes	yes	yes
Solvent Resistance	poor	good	good	fair	poor	good	good

spiral, and capillary (hollow fine fiber). These are illustrated in Figure 3-2.

In the tubular configuration, the membrane is cast on the inside of a 1/2 to 1 inch diameter porous support tube. The water to be treated flows through the center of the tube, and the permeate passes through the membrane and support tube into a collection vessel. These configurations have a very wide range of operating pressures, and good resistance to plugging by suspended solids due to the high flow velocity. The permeate production per unit volume is however very poor compared to the spiral wound or capillary units. Thus they are not cost competitive at low suspended solids levels, but can be used on streams with large amounts of suspended solids without the pretreatment required for spiral or capillary units.

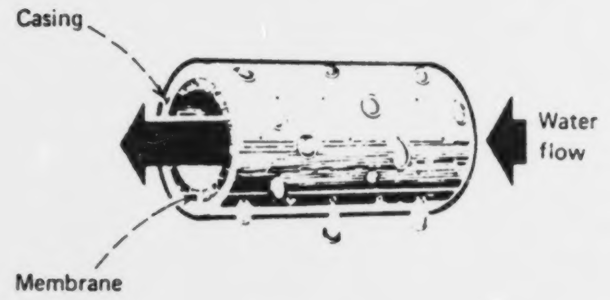
The spiral wound or scroll type membrane has a configuration which provides a high membrane surface area per unit volume ratio. The manufacturing technique involves wrapping a spiral sandwich of two sheets of membrane epoxy sealed along their common edge, and separated by a porous synthetic backing material which forms a channel through which the permeate flows. The permeate travels spirally into a central perforated permeate collector tube running longitudinally up the centre of the module. A polypropylene turbulence promoter acts as a spacer between the spiral roll of the membrane module.

In capillary, or hollow fine fiber configurations, the membranes are spun into hollow fibers which are very fine (100 to 200 micron in diameter) with walls about 25 micron thick. Cylinders with such low diameter do not need a supporting structure to withstand high pressures. Bundles of these long fibers are arranged in a U-shape,

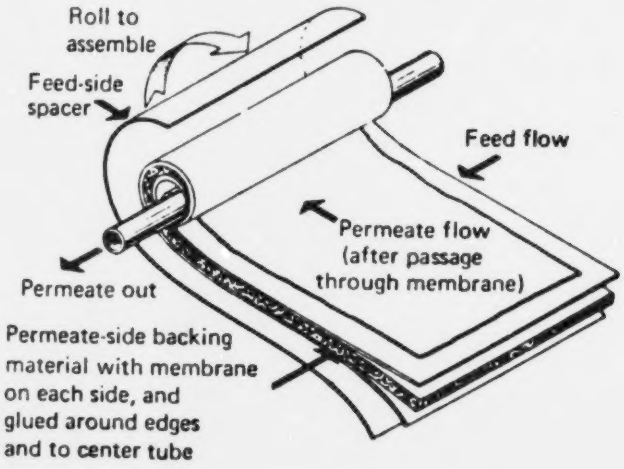
FIGURE 3 - 2

MEMBRANE CONFIGURATIONS

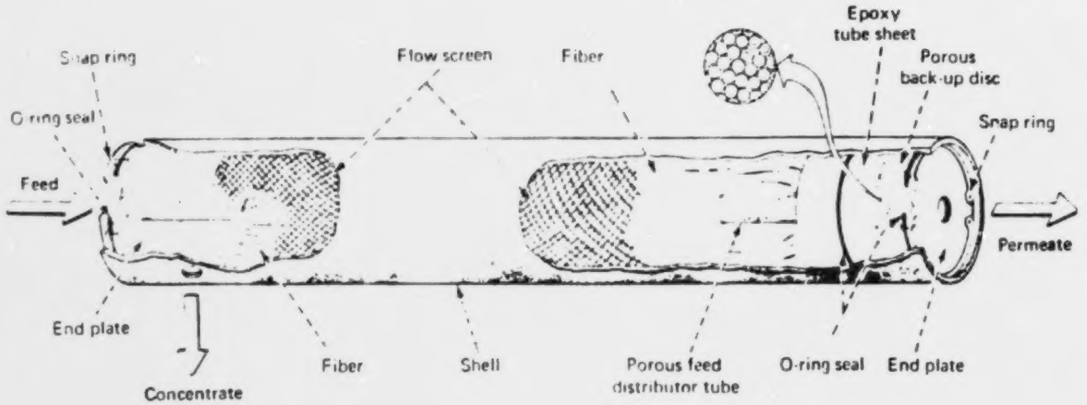
REPRESENTATION OF A TUBULAR MEMBRANE


Casing
Water flow
Membrane

FABRICATION OF A SPIRAL-WOUND MODULE


Roll to assemble
Feed-side spacer
Feed flow
Permeate flow (after passage through membrane)
Permeate out
Permeate-side backing material with membrane on each side, and glued around edges and to center tube

CAPILLARY(Hollow Fine Fiber) PERMEATOR


Snap ring
O-ring seal
Feed
End plate
Concentrate
Fiber
Flow screen
Shell
Fiber
Porous feed distributor tube
Epoxy tube sheet
Porous back-up disc
O-ring seal
Snap ring
End plate
Permeate

and the open ends are "potted" in plastic. This bundle is arranged in a cylinder vessel as shown in Figure 3-2. The feed solution is introduced into the middle of the fibre bundle through a porous tube at one end of the cylinder, and the concentrated product leaves from an effluent tube at the same end. The permeate passes through each fiber to the "potted" end where it is collected. Table 3-4 summarizes relative advantages and disadvantages, along with general pretreatment requirements for each configuration.

Usually configurations are selected based mainly on economical considerations, and spiral configurations, due to the large membrane surface areas, are often preferred for large scale systems. However, the configuration selected does determine pretreatment requirements. All configurations require removal of free oil and larger particulate to some extent. Tubular configurations have the least stringent pretreatment requirements and only require removal of particulate matter larger than 50 microns.

Spiral configurations have a high membrane surface area, and are often preferred for larger-scale applications. However, they generally require removal of particles larger than 5 micron in size and are more dependant on pH, since chemical glues and adhensives are used in the system housing which may be affected by pH, while tubular systmes use mechanical sealing components such as O'rings.

Hollow-fibers have similar limitations as spiral, although have not yet been widely applied in ultrafiltration.

TABLE 3-4

COMPARISON OF MEMBRANE CONFIGURATIONS

Configuration	Advantages	Disadvantages	Status	Pretreatment
Tubular (conventional)	Easily cleaned mechanically or chemically; very tolerant to suspended solids. Much operating experience Individual tubes replaceable	High holdup per unit membrane area Relatively expensive Requires moderately large floor space per sq.ft. although tube modules can be placed separately in and around existing equipment in special cases	Commercial	P, 50
Spiral-wound	Low in cost Compact; low floor space per sq.ft. Low hold up per unit membrane area Much operating experience	Easily plugged Cleaning options are limited	Commercial	P, 5
Hollow-fiber(shell-side feed)	Very low in cost Very compact Low holdup	Plugs easily Very hard to clean	Commercial	P, 5
Hollow-fiber (feed inside fibers)	Very low in cost Very compact Low holdup May be amenable to cleaning	Just now on market-meager operating experience	Fiber modules new on market	P, 5

P = pH Adjustment
5 = 5 micron particle size prefiltration to prevent clogging of the membrane channel of feed flow.
50 = 50 mesh filter to remove particles which would damage pumps.

3.2.2 Current Applications and Economic Evaluation

There is a wide range of applications in industry where ultrafiltration has been proven and is being proven to be a viable alternative for pollution control and product recovery (Lopez and Johnson, 1977; Christensen and Plauman, 1981; Koehler, 1983).

Ultrafiltration has been utilized in treatment of emulsified oily wastes from machining operations, automotive plants, and rolling mill coolants, for either the recovery of oil or production of water suitable for reuse. Due to the general lack of research on ultrafiltration processes specific to oilfield applications, it is necessary to examine efficiency and economics of well-documented applications in other related industries. Typically industrial oil/water emulsions exist at a concentration of 0.5-10% oil in the feed and UF can remove 90-95% of the water in the emulsion up to about 30-40% oil (Dhawan, 1978).

Both cellulosic and non-cellulosic membranes have been successfully employed in industrial applications, depending on feedwater characteristics and treatment requirements. Cellulose acetate membranes have been sucessfully employed in some cases, such as for treatment of cutting oil from the metal working industry (Kutowy et al, 1981), although their susceptibility to hydrolysis at high and low pH has been a limiting factor.

Early research by Goldsmith, et al, (1974) studied the treatment of soluble oil waste from an electric company, which contained 1 to 3 per cent oil, of which 30% was emulsified machine coolants. They used one inch tubular polymeric membranes, and achieved a concentrate oil level of 40 percent. Fouling was not encountered in this study, and this was attributed to the presence of detergent emulsifiers which prevented an oil build-up on the membrane.

The efficiency of the treatment of stable oil emulsions from the metal working industry by ultrafiltration was reported by Dick (1981). The use of non-cellulosic hollow fine fiber ultrafiltration was found to economically remove 98% of oil with 90% reduction in volume. The oil in the UF concentrate was suitable for further treatment to recover the oil.

A survey of current tubular ultrafiltration systems in use for oily wastewater treatment in various industries (Wahl et al, 1979) found plants as large as 100,000 gpd were successfully and economically operating, and the resulting oil recovery in some cases resulted in breakeven, or even profitable operations, as shown in Table 3-5.

In addition to the well documented efficiency of UF, combinations of UF with subsequent Reverse Osmosis RO application on UF permeate have been used on oily wastewaters for oil recovery, and to produce a water quality suitable for industrial reuse. Mixed industrial wastes containing process wastes such as cutting oils as well as rinse tank waters, cooling tower bleeds, and miscellaneous other wastes, have been treated by UF/RO combinations to provide water quality suitable for direct recycle (Lopez and Johnston, 1977). Reverse Osmosis will be discussed in greater detail in section 3.3.

A combination of UF/RO was used for treatment of oily waste streams in the aluminum industry (Sonksen, 1978). Alcoa investigated and compared the operation and economics of chemical, evaporation, UF, and UF/RO treatment of their wastes. Pilot plant studies concluded

TABLE 3 - 5

SURVEY RESULTS - OILY WASTEWATER TREATMENT SYSTEM ANNUAL COST SUMMARY

(from Wahl et al, 1979)

System	Date of System Start-Up (mo/yr)	Design Flow Rate (gpd)	Actual Daily Flow (gpd)	Direct Labor ($)	Power ($)	Membrane Replacement ($)	Cleaning Chemicals ($)	Concentrate Disposal (Oil Recovery Credit) ($)	Total Cost (Ex. Oil Credit) ($)	Cost Per Gallon (¢)	Total Cost with Oil Recovery Credit (Savings) ($)
1	5/76	18,000	20,000	20,000	6,893	27,653	17,386	(150,000)	71,932	1.4	(78,068)
2	7/77	1,000	800	8,000	895	1,298	653	4,400	15,246	7.6	–
3	12/74	2,500	4,300	3,125	2,954	1,540	6,224	(91,000)	13,843	1.3	(77,157)
4	2/78	1,000	1,100	6,250	1,253	3,060	840	(325)	11,403	4.1	11,078
5	9/75	1,000	500	6,667	895	325	2,000	0	9,887	7.9	
6	7/75	2,500	4,000	5,500	985	0	2,425	0	8,910	0.9	
7	9/75	10,000	6,000	2,857	2,686	139	2,415	(8,550)	8,097	0.5	(453)
8	9/77	3,000	2,000	125	985	0	–	–	–	–	
9	11/75	5,000	5,000	3,000	2,954	2,720	5,510	—	–	–	
10	11/75	8,000	4,000	2,917	2,686	0	277	1,400	7,280	0.7	
11	3/77	4,000	10,000	5,625	2,954	-0	1,750	(9,375)	10,329	0.4	954
12	9/75	20,000	40,000	7,000	6,983	3,338	100	(64,000)	17,421	0.2	(46,579)
13	11/74	4,000	4,000	8,750	2,328	1,322	1,081	(68,640)	13,481	1.3	(55,159)
14	5/76	3,000	3,000	9,286	1,432	1,190	648	3,840	16,396	2.2	
15	6/77	100,000	100,000	20,000	24,170	3,326	17,051	(79,688)	64,547	0.3	(15,141)
16	3/75	1,728	3,000	3,611	985	1,265	200	0	6,061	0.8	
17	12/74	500	500	1,548	985	1,078	48	2,160	5,819	4.7	
18	2/76	15,000	20,000	11,000	5,640	6,277	5,533	(1,700)	28,450	0.6	26,750
19	9/75	980	300	250	895	325	48	30	1,548	2.1	
20	3/75	500	1,000	5,833	985	428	333	–	7,579	3.0	
21	11/74	500	300	3,125	895	2,116	250	–		–	

that UF/RO would provide a superior and reusable product than the chemical treatment system currently employed. Estimated treatment costs (1978 dollars) for a 100,000 gpd plant are presented in Table 3-6. The figures show a slightly lower capital investment for the membrane processes. After one year's operation on full scale, actual operating costs were $0.76/1,000 gal lower for chemical treatment, but this did not include cost of sludge disposal or possible post treatment of oil to meet regulatory effluent guidelines. Problems were encountered with membrane fouling and associated cleaning requirements, but operators still were most satisfied with the combined UF/RO facility based on product quality, lower sludge generation, and lower dependence on operator judgement.

Ultrafiltration has also been shown to be efficient and economical for pretreatment of seawater prior to desalination by reverse osmosis (Hayward and Choate, 1982). The ultrafiltration process was used to retain suspended, emulsified, and colloidal solids, as well as dissolved macromolecules, and produce a permeate which resulted in stable RO productivity and little maintenance or operator attention. The purpose was to prevent fouling, and thereby increase membrane life. After 9,600 hours of operation the RO permeate had not yet required cleaning or regeneration. This pretreatment is now thought be technically excellent and economically competetive with conventional pretreatment, thus illustrating the potential of UF not only for oil removal but also as a pretreatment to RO on waters with high levels of suspended solids or dissolved macromolecules.

TABLE 3 - 6

a) ESTIMATED COST ANALYSES AT ALCOA PENNSYLVANIA

(Sonksen et al, 1978)

System	Capital Cost	Total Unit Operating Cost ($/1000 gal)	Net Unit Operating Cost ($/1000 gal)
Chemical	$1,730,000	$11.33	$5.36
Evaporation	2,650,000	8.75	2.90
UF	1,450,000	7.73	1.77
UF/RO	1,550,000	8.81	1.12

b) OPERATING COST COMPARISONS AT ALCOA

	Chemical	Ultrafiltration	Ultrafiltration and Reverse Osmosis
Depreciation	$2.58/1,000 gal	$2.44/1,000 gal	$2.69/1,000 gal
Power	0.12	0.47	0.62
Steam	0.10	0.19	0.19
Labor	1.73	2.00	2.00
Chemicals	2.00	1.48[a]	1.48[a]
R & M	1.54	0.92	0.92
Membrane replacement	--	1.61[b]	2.30[b]
Total unit operating cost	$8.07/1,000 gal	$9.11/1,000 gal	$10.20/1,000 gal
Credit for water recovery	--	$0.05/1,000 gal	$ 1.37/1,000 gal[c]
Projected oil recovery credit[d]	$5.30/1,000 gal	$5.30/1,000 gal	$ 5.30/1,000 gal
Projected net operating cost	$2.77/1,000 gal	$3.76/1,000 gal	$ 3.53/1,000 gal

[a] This unit chemical cost is based on cleaning the membrane after every 100,000 gal batch.

[b] Membrane replacement costs are based on a two year membrane life.

[c] Permeate from the reverse osmosis system will be used as feed to the deionized water system. Because of its superior quality (<50 μmhos conductivity), the length between resin regenerations will be significantly extended. This credit represents the savings on chemical and labor costs over a full year.

[d] Projected oil recovery credits based on 80% recovery of the incoming oil and 20¢/gal net value.

Capital and operating costs of ultrafiltration systems are strongly dependent on the specific application, capacity, and type of system. The literature has identified many applications where UF has been effectively and economically applied. General economic advantages of systems have been given as lower chemical costs, low pretreatment costs (depending on configuration), lower energy costs since the main energy input required is to operate pumps, and lower labour requirements for operation. As well, there are economic benefits from oil recovery. The main operating costs of UF are associated with membrane replacement and cleaning.

3.2.3 Applications in the Oilfield

Studies examining the application of ultrafiltration systems to the treatment of oily wastewater associated with oil production have been limited to date. Contact with various research agencies, and the literature review revealed that little research is being conducted in this area. Generally, the relatively small size of most UF systems in current use, lack of applications experience and problems identified with membrane fouling have been factors discouraging oilfield applications. The Alberta Oil Sands Technology and Research Authority (AOSTRA) has undertaken studies examining the use of ultrafiltration for oilfield produced waters, however results have not been published in the open literature. Energy, Mines and Resources Canada has undertaken research investigating ultrafiltration of in-situ generated bitumen emulsions. The feasibility of using porous membranes to dewater and break bitumen oil/water emulsions was shown by Farnand and Poirier (1983). They tested several membranes and found that the emulsion apeared to break on the surface of the membrane. Cellulose membranes were found to foul the least and were considered most attractive due to their hydrophilic nature. The study identified the potential suitability of other hydrophilic ultrafiltration membranes for use in treating such emulsions. The study found, however, that static test cells exhibited low permeation rates due to a build up of a polarized layer on the surface of the cell.

The static studies were followed up by tests using a continuous process (Farnand, Coulombe, and Sawatsky, 1984, and Nader, Farnand and Sawatsky, 1984). They confirmed that the use of a continuous membrane process to break these emulsions is feasible and they identified a polar material in the permeate which was considered to be the stabilizing agent. The main concern was identified to be membrane fouling, which was related to membrane material used.

Ultrafiltration has been examined for treatment of emulsions from simulated micellar-flood wastewaters by the U.S. Department of Energy (Dubey et al, 1981). These were complex wastewaters containing oil, as well as surfactants, alcohols, and polymers, that were introduced to the formation during micellar flooding. Synthetic wastewaters were prepared to simulate micellar flood waters, containing fixed levels of brine, polymer and butanol, with oil-sulphonate concentrations ranging from 400-3400 mg/L. Operating pressures were 20-40 psi (138-276 kPa) and temperatures ranged from 21-32°C (70-90 °F). Depending on oil concentration, flux rates of cellulosic membranes ranged from 20-40 gal/ft^2/day after 180 min, and the permeate was essentially free of all oils, sulphonate, and polymer. Polymeric membranes were also tested, and the researchers found that both cellulosic and polymeric membranes produced permeates relatively free of oil sulphonate or polymer, although the cellulosic was said to provide better flux rate and was easier to clean when fouled. This study also identified the potential of ultrafiltration for treating produced waters which are destined for subsurface disposal or reinjection, since significant amounts of emulsified oil could result in formation plugging and the associated production problems. The study concluded that ultrafiltration was effective, but that further studies on actual wastewaters were necessary due to their complex and site specific characteristics.

Ultrafiltration has also been investigated for recovery of surfactant from oilfield emulsions (Baddour, 1977). The patent claimed that further concentrating ultrafiltration permeates and concentrates by membrane processes resulted in an efficient and economical process for recovering surfactant for subsequent reuse in the oilfield.

The literature has shown the wide range of UF industrial applications which have proven its technical and economical efficiency. Operating experience has shown that the advantages of UF, as compared to other conventional chemical and mechanical processes, are the effective concentration and recovery of oil, elimination of oily sludge disposed problems, a consistent and high oil removal efficiency which is virtually unaffected by feed concentration (up to the point where the emulsion breaks), and the low pretreatment and chemical requirements.

In addition to removal and concentration of oil as a step in boiler feedwater preparation, ultrafiltration may be utilized in several other applications in the oilfield. UF may be used as a polishing step in conventional boiler feedwater treatment processes for removal of residual levels of suspended solids, turbidity and oil. As well, UF may economically provide the less rigorous treatment necessary prior to deep well disposal of produced waters to prevent formation clogging. Research has also shown that surfactants may be recovered through concentration by membranes and reinjected for recovery operations.

Ultrafiltration permeate may also be further polished by Reverse Osmosis in order to meet boiler feedwater requirements, as discussed in the following section.

3.3 Preparation of Boiler Feedwater by Reverse Osmosis

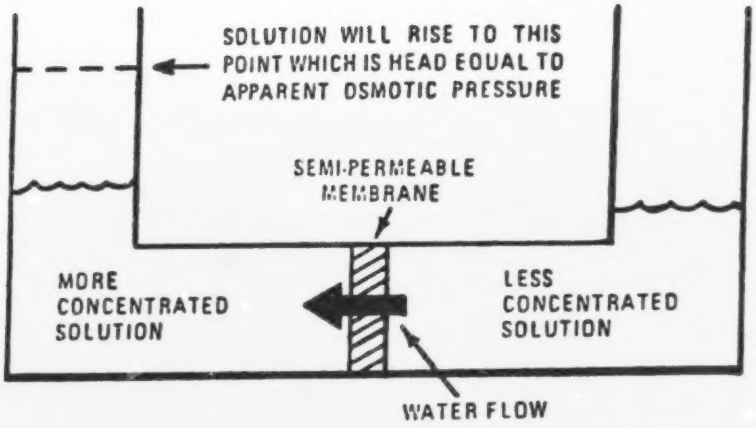
3.3.1 Process Description

As shown in Figure 3-3, osmosis is the spontaneous passage of water through a semi-permeable membrane from a less concentrated solution to more concentrated solution until equilibrium is reached. The liquid head developed at the equilibrium point is defined as the "Osmotic Pressure" (about 1 psi or 6.9 kPa per 100 mg/L total dissolved solids for water). Reverse Osmosis occurs when an applied pressure exceeding the Osmotic pressure is applied on the concentrated solution and the passage of pure water is reversed travelling from the more concentrated solution to the less concentrated solution.

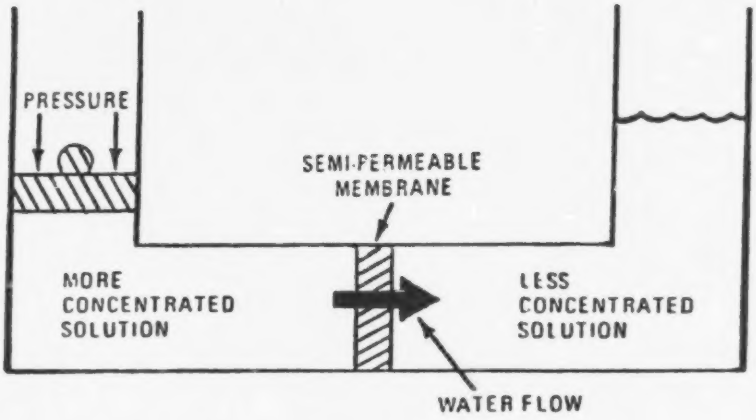
In reverse osmosis systems, the feedwater is passed across the membrane surface under turbulent flow and high pressure conditions. As the stream flows across the membrane, pressure forces pure water through the membrane. The impurities in the feedwater are

FIGURE 3 - 3

REVERSE OSMOSIS


SOLUTION WILL RISE TO THIS POINT WHICH IS HEAD EQUAL TO APPARENT OSMOTIC PRESSURE
SEMI-PERMEABLE MEMBRANE
MORE CONCENTRATED SOLUTION
LESS CONCENTRATED SOLUTION
WATER FLOW

OSMOSIS


PRESSURE
SEMI-PERMEABLE MEMBRANE
MORE CONCENTRATED SOLUTION
LESS CONCENTRATED SOLUTION
WATER FLOW

REVERSE OSMOSIS

concentrated and carried out of the unit. Reverse Osmosis employs an anisotropic membrane with an asymmetric structure consisting of a 0.1 to 0.5 micron thick dense layer (active surface), supported by a 25 to 100 micron thick substructure.

Reverse osmosis systems are operated at a higher pressure than UF. Membrane pore sizes are smaller (5 - 50 Å), and the two mechanisms of separation are ionic salt separation and organic separation. Ionic salt separation is due to the dielectric interaction and interfacial tension effects at the water-membrane interface. The higher the charge the greater the rejection or repulsion from the surface layer. As a result of this repulsion a pure water layer sets up on the membrane surface and extends into the pore structure of the membrane. Divalent ions are rejected to a greater extent (96-99%) than monovalent ions (94-96%).

Organic separation is based primarily on the shape and size of the molecule and its relationship to location at pore sites since no pure water layer is established on the membrane surface. Extensive research has been conducted investigating the separation of various organic compounds by membrane processes (Matsuusa and Sourirajan, 1973 a and 1973 b; Dickson et al, 1975). Generally, organics with a molecular weight greater than 300 are rejected almost completely, with typical percentage rejections typically ranging from 90-99.9%, depending on the charge, molecular size, interaction forces, configuration and degree of dissociation.

In designing Reverse Osmosis systems, the feed water source and the final desired water quality are the main criteria in selecting the type of system for a particular application. Reverse Osmosis is a modular process, allowing for accurate scale-up from pilot processes.

Operating pressures can vary from 200 to 1000 psi (1380-6900 kPa) depending on the osmotic pressure of the feedwater. For example seawater with 14,000 mg/L of total dissolved solids (TDS) has an osmotic pressure of approximately 350 psi (2415 kPa) and city water, with 200 mg/L of TDS has an osmotic pressure of about 5 psi (35 kPa). The effective applied pressure is the pressure applied minus the osmotic pressure. For seawater and brackish water, the osmotic pressure plays an important role in determining the operating pressure. In processing of typical surface water the osmotic pressure is only a small factor.

3.3.1.1 Membrane Materials

The selection of a reverse osmosis membrane system to treat produced water is based primarily on the composition of the de-oiled feedwater and the quality of the product water required. Boiler feedwater treatment requirements were described in section 2.2, and these criteria set the objectives for development of the reverse osmosis membrane process for this application.

Table 3-7 presents a summary of commercially available reverse osmosis membranes. These may be classified as cellulosic and non-cellulosic membranes. Non-cellulosic membranes give better organic rejection than cellulosic membranes. However generally, cellulosic membranes are less expensive and are more productive in terms of flux rates. The most commonly used membranes are cellulose acetate and thin-film-composite.

TABLE 3-7

COMMERCIAL REVERSE OSMOSIS MEMBRANES

Membrane Material	Configuration	Salt Separation (%)	Permeate Rate ($Lm^{-2}d^{-1}$)	Pressure (kPa)	pH (range)	Max. Temp. °C
Cellulose Acetate	Spiral	90-99	300-1000	2900-7200	4-7	25-35
	Capillary	97	200	2900	6-8	35
	Flat	<99	400	4000	2-8	30-40
	Tubular	90-98	450-700	4000	3-8	30-45
Co-Polyacrylonitrile	Flat/Tubular	95-98	1000-3000	5000-10000	1-10	45
LP-Polyamide	Spiral	96	1000	1400	2-11	35
Piperazine/ Polyepiiodohydrin/ Isophthaloyl Chloride	Spiral	>99	700	4000	4-12	40
Polyamide	Capillary	95-98.5	40-50	2900-5800	4-11	35
Polyamide TFC	Spiral	96-98	750	7200	2-10	45
Polybenzimidazolone	Spiral	>97	1000	10000	1-12	60
p-Phenylenediamine/ Trimesoyl Chloride	Flat	98	1800	4100	2-11	<60
Polysulfone, sulfonated	Capillary	N/A	N/A	N/A	2-13	70

(adopted from Pusch and Walch, 1982)

Thin-film-composite (TFC) reverse osmosis membranes are composed of two separate layers. The first layer is a thin barrier layer which performs the water and salt separation and may be formed from any of a number of membrane materials. The second layer is a microporous sublayer which supports the barrier layer. Both conventional and thinfilm composite membranes also have a third layer, a fibrous web, which serves as a back-up for the microporous support.

For each type of membrane selected there will be associated operating limitations; most Reverse Osmosis membranes have operating temperature limits of approximately 35-40°C (95-104°F) although some have been developed for conditions of up to 60°C. As well, cellulose acetate membranes have a relatively narrow pH range, whereas non-cellulosics are capable of operation at a much wider range of pH. Applied pressures also vary; for example the maximum pressure for cellulose acetate, polyamide, and TFC are 500 psi, 800 psi, and 1000 psi respectively.

The literature has revealed extensive variability in the concentration of inorganic species in produced water, therefore preliminary testing must be done in order to evaluate membrane suitability, and to determine associated operating conditions, and required pressures.

3.3.1.2 Configurations

Various configurations of Reverse Osmosis membranes are available, and as in ultrafiltration applications, are selected based on economics and operating conditions which are determined by feedwater characteristics. RO membranes are available in spiral, tubular, and capillary configurations similar to these which were described in detail in section 3.2.1.2 for ultrafiltration, and similar operating limitations apply. In contrast to ultrafiltration, there is a wide body of experience in the operation of large-scale RO capillary systems.

For RO systems, additional pretreatment is often required to soften the feedwater. Depending on feedwater hardness levels, this may be necessary to prevent precipitation or scale formation on the membranes.

3.3.2 Current Applications and Economic Evaluation

Reverse osmosis has been sucessfully used in treating a variety of waters and wastewaters for industrial and municipal use, and has been accepted as a pretreatment step for demineralization, as well as for recycling process water at zero discharge plants (Strauss, 1981). Large scale reverse osmosis plants treating up to several million gallons per day are currently operational.

Reverse osmosis has been used to demineralize a variety of waters from low salinity municipal water to high salinity brackish and seawater to produce water for drinking, boiler feed, industrial, pharmaceutical, and medical uses. Pretreatment is often required to lower the pH (to less than 6.5 with cellulose acetate membranes) and reduce suspended solids for certain configurations. Post treatment with a second stage RO or demineralizers to reduce the soluble ions to the required levels may also be implemented, depending on product water requirements.

Reverse osmosis has proven to be efficient and economical for treatment of metal finishing wastewaters (Franza, 1981). The systems are used for purification of process waters for recycle, and recovery and reuse of the chemical concentrate. The application of membrane technology to power generation waste was discussed by Tang and Chu, (1981) and Dhwan and Doiel (1983), and considered efficient and economical. Total dissolved solids in power plant wastes were reported by Dhwan and Doiel (1983) to have been reduced by 95%, and silica reduction was in the order of 98%.

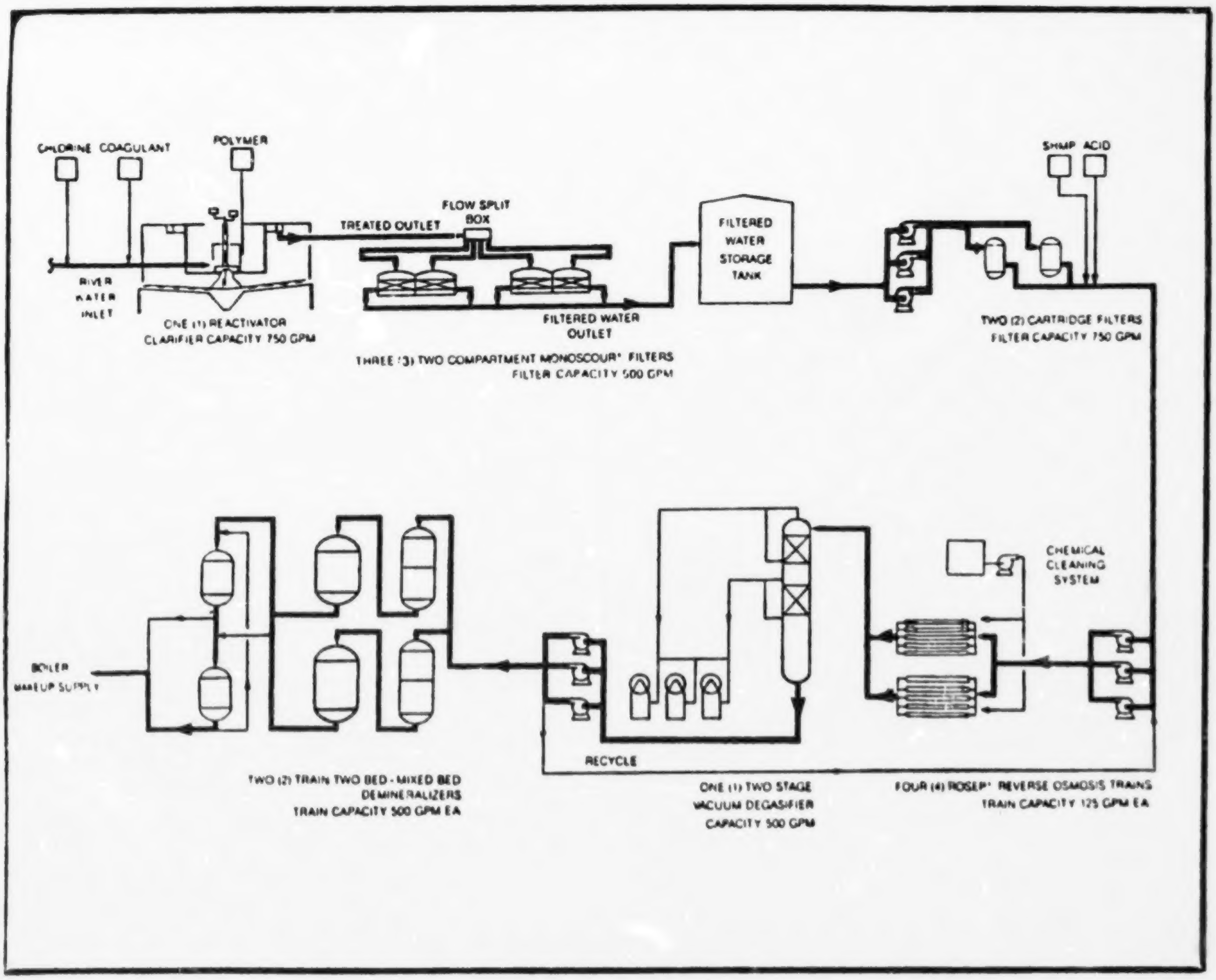
Reverse osmosis has been used in treating various waters for boiler feed water use in combination with some pre-treatment and post-treatment (Anonymous, 1976). O'Brien and LaTerra, (1981) reported on operating experience with six large reverse osmosis plants ranging in size from 450,000 GPD to several million GPD. Five of the case histories utilized the process for utility boiler make-up systems, while one was producing ultra-pure water. The feed contained total dissolved solids levels ranging from one hundred mg/L to as high as several thousand mg/L. One system was successfully handling 700,000 gallons per day of feedwater containing up to 5,000 mg/L TDS from a tidal river source. A spiral wound cellulose acetate system was with final demineralization by ion exchange. The treatment schematic is illustrated in Figure 3-4 . Projected salt rejection after 3 years operation was 91.9%.

In another system, regulatory guidelines required that a utility was to find an alternate water source to the well it was currently using. The high TDS feedwater levels (2,000 mg/L) obtained from the alternative man-made reservoir were too high to utilize in

FIGURE 3 - 4

PROCESS SCHEMATIC FOR RO TREATMENT SYSTEM ON HIGH TDS FEEDWATER

(from O'Brian and LaTerra, 1981)


CHLORINE
COAGULANT
POLYMER
RIVER WATER INLET
ONE (1) REACTIVATOR CLARIFIER CAPACITY 750 GPM
TREATED OUTLET
FLOW SPLIT BOX
FILTERED WATER OUTLET
THREE (3) TWO COMPARTMENT MONOSCOUR* FILTERS FILTER CAPACITY 500 GPM
FILTERED WATER STORAGE TANK
SHMP
ACID
TWO (2) CARTRIDGE FILTERS FILTER CAPACITY 750 GPM
CHEMICAL CLEANING SYSTEM
FOUR (4) ROSEP* REVERSE OSMOSIS TRAINS TRAIN CAPACITY 125 GPM EA
RECYCLE
ONE (1) TWO STAGE VACUUM DEGASIFIER CAPACITY 500 GPM
TWO (2) TRAIN TWO BED - MIXED BED DEMINERALIZERS TRAIN CAPACITY 500 GPM EA
BOILER MAKEUP SUPPLY

the existing ion exchange plant. For the spiral wound cellulose acetate system pre-treatment consisted of clarification and filtration, and optional post treatment demineralization. After six months operations the plant was still rejecting 96% of the salt, and just experiencing a first flux decline.

Another large plant treating over 1 MGD of high TDS water for boiler make-up utilizes clarification and filtration as pre-treatment and demineralization post treatment similar to that shown in Figure 3-4. This plant identified the operational advantages of adding a scale inhibitor, such as sodium hexametaphosphate to the system. In summary, the plants examined in the study were operating reliably and cost-effectively, and potential for further expansion and application of the technology was forseen.

The ability of Reverse Osmosis to handle extremely high levels of dissolved solids has been illustrated by various installations converting seawater to potable water or for boiler feed. An example is the National Electric Company of Venezuela, where a two-stage RO has been successfully treating seawaters containing an average of 36,000 mg/L of T.D.S. (Vera, 1981). Pre-treatment was required to remove suspended solids and organic matter to protect membrane modules. The capital cost of pre-treatment constituted approximately 18% of the total system capital cost, with the Reverse Osmosis System constituting 56%. Other costs were associated with piping controls and electrical materials. The RO process was found to increase the length of time between regenerations of an ion exchange demineralization unit, through which RO product water is processed prior to entering hign pressure (2,500 psi or 17250 kPa) boilers.

One of the world's largest seawater desalting facilities, utilizing reverse osmosis to produce water for steam generators for enhanced oil recovery was constructed in 1979. The system consisted of nine RO units with a total capacity of 3.6 million gallons per day and was constructed to provide feedwater for generators in steamflood operations adjacent to the Caspian sea (Leitner, 1979). Although operating data was not provided the hollow fine fiber systems were expected to remove 98.5% of dissolved solids from the seawater. The system was compared to distillation technology, and considered to be more economical due to low energy needs and lower maintenance costs. The pre-treatment utilized was acid, polyelectrolyte, and bisulfite addition, followed by sand filtration, sodium hexametaphosphate addition, and cartridge filtration. Post-treatment was not required.

The literature has illustrated the well documented technical efficiency of reverse osmosis for treatment of waters containing extremely high levels of TDS, as well as for removal of organic compounds, and the resultant production of a water quality suitable for boiler feedwater, process recycling, or potable use. The cost of RO systems are difficult to compare due to the extreme variability in feedwater quality and associated operating pressures which are required.

The capital cost for RO is also dependent on the quality of feedwater, size of system, pre and post-treatment, and membrane configuration. As with UF systems tubular configurations are often several more times expensive than spiral systems as a result of the lower membrane surface area to system volume ratio. RO has been implemented in numerous large scale industrial applications treating up to several million gallons per day. The economic advantage of RO have been credited as their mechanical simplicity, low energy requirements and low chemical and maintenance cost. As in UF, the main operating costs are membrane replacement and cleaning.

3.3.3 Applications in the Oilfield

The potential for the application of reverse osmosis for boiler feedwater preparation has been identified by the oil industry. Reverse osmosis has been evaluated for a limited number of oilfield produced water applications, although systems have not yet been commissioned.

Previous discussion in section 2.3.2 showed that at Cold Lake, reverse osmosis was one system evaluated. However, economics at that time favoured the conventional processes. AOSTRA has sponsored some research into the application of reverse osmosis for treatment of in-situ produced water, but results have not been published and are unavailable to-date. One published study by the U.S. Department of Energy (Hicks and Liang, 1981) was conducted to determine the feasibility of reverse osmosis for treatment of oil shale processing wastewater steams. Although water composition varies from that of in-situ/heavy oil operations, especially with regard to organic composition in retort waters, some waste streams are directed to steam generation and similar treatment objectives were required. The study found that reverse osmosis was successful in treating retort waters to a water quality suitable for reuse within the plant, with TDS and organic reductions from 93-99%.

Reverse osmosis has shown to be economically competitive with the more conventional systems utilized in industrial processes. Advantages of RO have been cited as lower operating and chemical costs, lower energy requirements since the only energy required is to run the pumps, and mechanical simplicity. It is now a well established technology and large systems treating several million gallons per day are common. RO has also been utilized prior to ion exchange and has served to increase the life of the resins. It also removes dissolved organics that could normally foul the resin.

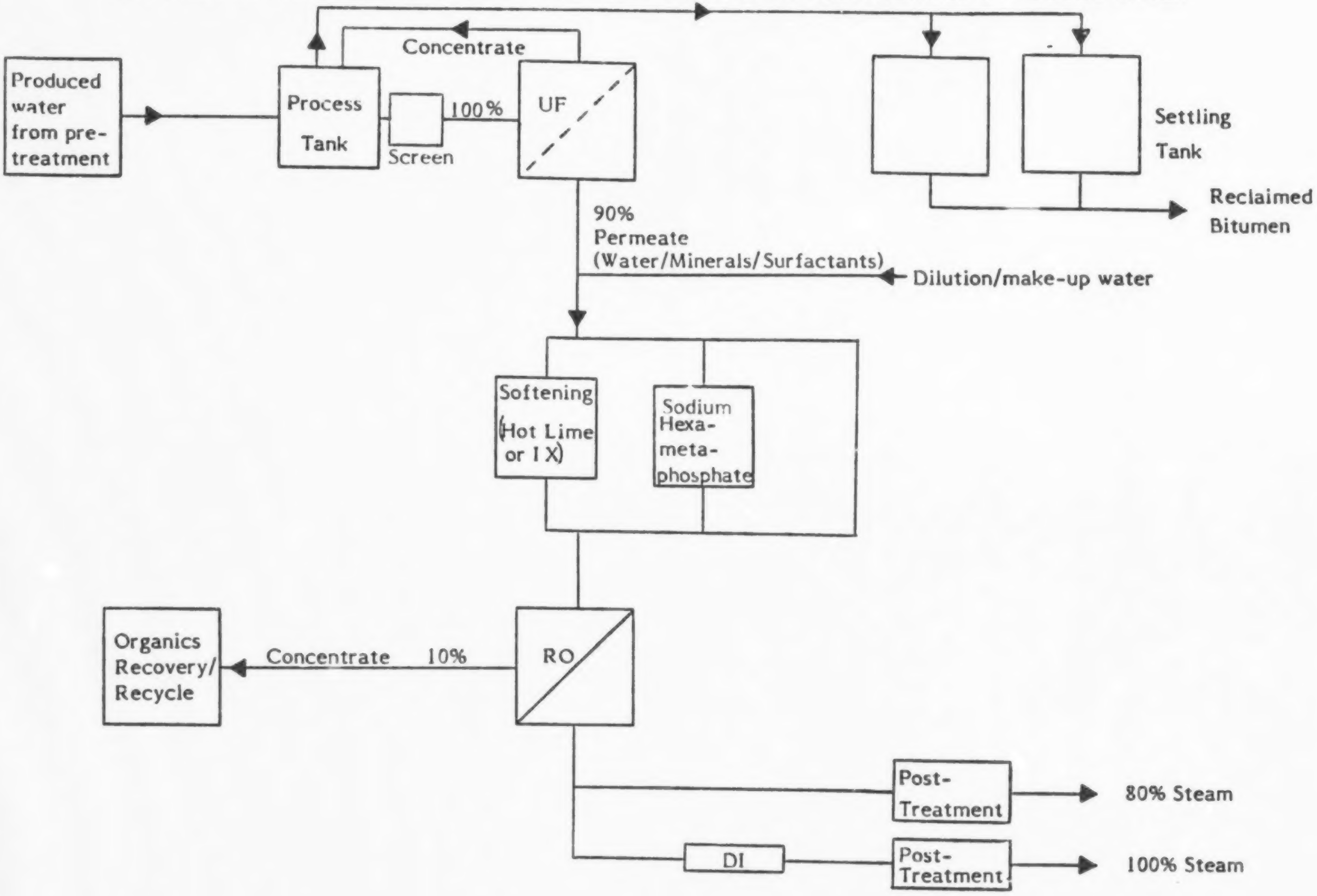
The potential for reverse osmosis in oilfield applications for boiler feedwater treatment has therefore been illustrated, however, little published information exists on actual laboratory or field evaluation. Most RO systems have been installed primarily to reduce total dissolved solids levels. Since the main treatment concern in recycling produced water for use as boiler feed water is usually the reduction of hardness levels, the potentional of RO has often been overlooked. The use of membrane technologies for treatment and recycling of produced waters has the ability to not only reduce hardness, but may result in a water quality better than that required by conventional oil-field steam generators, thus offering the potential for assessment of alternative and more efficient steam generation technologies.

3.4 Potential UF/RO Membrane System Design

Based on the literature review and related applications several schematics using membrane technology to treat produced water for steam generation can be envisaged. Figure 3-5 shows the type of

FIGURE 3 - 5.

PRE AND POST TREATMENT OPTIONS ON A UF/RO SYSTEM FOR TREATMENT OF PRODUCED WATER


Produced water from pre-treatment
Process Tank
Concentrate
Screen
100%
UF
Settling Tank
Reclaimed Bitumen
90%
Permeate
(Water/Minerals/Surfactants)
Dilution/make-up water
Softening (Hot Lime or I X)
Sodium Hexa-meta-phosphate
RO
Concentrate
10%
Organics Recovery/ Recycle
Post-Treatment
80% Steam
DI
Post-Treatment
100% Steam

possible pre and post treatment processes that may be employed to meet oil recovery and boiler feedwater objectives.

The design of a treatment system employing UF/RO processes is dependent upon numerous factors. These include feedwater composition, treatment requirements, and operating limitations of the membranes and configurations used. Prior to any UF system for oil removal, free or floating oil must be removed, usually by some form of gravity separation such as a skim or surge tank. Systems employing spiral configurations usually have stringent pre-treatment requirement, and may require filtration for the removal of all particulate matter greater than 5 microns in size. After primary oil removal, the feedwater would then flow to a process tank for flow equalization, and subsequently through the UF system. Concentrate could be recycled back to the process tank, and floating oil removed for bitumen recovery.

The addition of low-salinity, oil-free, make-up water between the UF and RO stages can also be envisioned for oilfield use to provide sufficient volume for boiler feeds. The diluted or undiluted ultrafiltration permeate could be pretreated by softening or pH adjustment if necessary prior to RO. If the RO permeate required further treatment to meet boiler feedwater objectives, ion exchange could be utilized. Post treatment includes oxygen scavenging and pH adjustment as required. Since such a system may well be capable of producing water quality better than that required for conventional steam generators, the potential exists for producing a quality suitable for 100% steam generators with deionization post treatment as necessary.

4.0 PROPOSED EXPERIMENTAL PROGRAM

The literature review has revealed the extensive variability in chemical characteristics of produced waters from heavy oil/in-situ bitumen operations. As a result treatability or evaluation test musts be conducted on specific waters to determine membrane process efficiencies. Testing determines the degree of species separation, product water quantity per square foot of membrane surface, and the recovery (ie ratio of product flow to feed flow) which can be obtained.

4.1 Recommended Membranes

Based on the discussions in previous sections and typical produced water quality as obtained from the literature, the following membranes are presented as the most promising for treatment of oil/water emulsions in heavy oil/in-situ produced water:

i) cellulose/cellulose acetate (C/CA)
ii) sulfonated polysulfone (SPS)
iii) polyvinyl difluoride (PVDF)
iv) thin film composite (TFC) (polyamide)
v) polysulfone (PS)

All of the above membranes are commercially available except for the cellulose membrane. Cellulose acetate has been proven in numerous oil/water applications, but has pH and temperature limitations and is prone to bacterial attack. Therefore the newly developed cellulose membranes, which have the advantage of a wider operating pH, will be evaluated. Both PVDF and the polyamide TFC are very hydrophilic and offer a very wide operating range of pH and temperature conditions. They are generally resistant to bacterial and solvent attack. The membranes selected above would allow testing at a wide range of pH and temperature conditions.

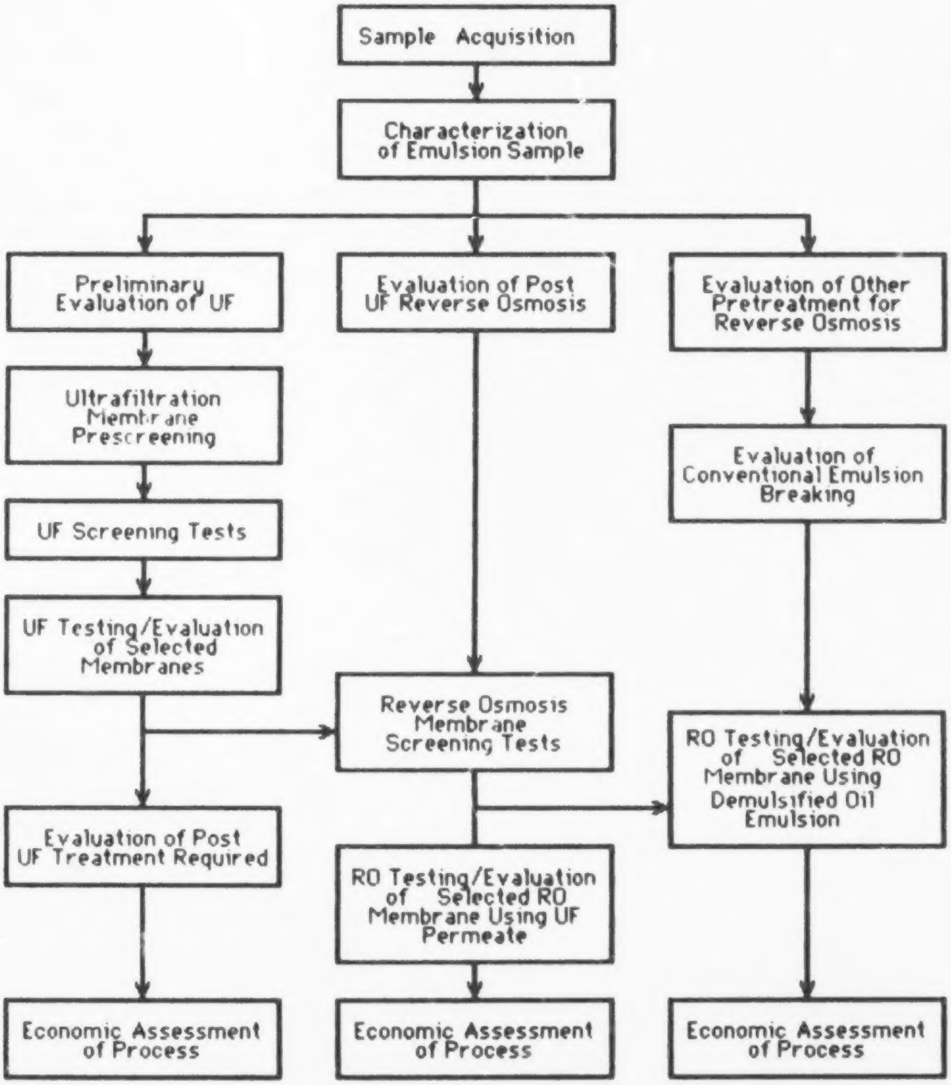
4.2 **Experimental Procedures**

A flow chart outlining the procedure for proposed work evaluating membranes and other treatment processes is shown as Figure 4-1. The analytical results from characterization of the produced water samples will assist in the determination of optimal UF membranes and membrane configurations as well as pretreatment requirements for UF. Membrane prescreening tests should be conducted to identify the optimum porosity of each membrane type being tested. Screening tests on optimum membranes will then identify the best membranes and operating conditions. Testing and evaluation of the best membranes should then be conducted to determine the quality of permeate and the flux rates which can be obtained under actual membrane system operating conditions. Post-treatment steps required for UF permeate to achieve boiler feed water quality should be assessed.

The analyses of ultrafiltration permeate will be used to determine the osmotic pressure and thereby assist in identifying the operating conditions and recovery limitations for subsequent reverse osmosis treatment. Reverse osmosis tests should be conducted using permeate from one of the selected UF membranes to identify the best RO membrane for use in this application and to evaluate permeate quality and flux rates under actual membrane operating conditions.

Figure 4 - 1

PROCEDURE FOR MEMBRANE AND PROCESS EVALUATIONS


Sample Acquisition
Characterization of Emulsion Sample
Preliminary Evaluation of UF
Evaluation of Post UF Reverse Osmosis
Evaluation of Other Pretreatment for Reverse Osmosis
Ultrafiltration Membrane Prescreening
Evaluation of Conventional Emulsion Breaking
UF Screening Tests
UF Testing/Evaluation of Selected Membranes
Reverse Osmosis Membrane Screening Tests
RO Testing/Evaluation of Selected RO Membrane Using Demulsified Oil Emulsion
Evaluation of Post UF Treatment Required
RO Testing/Evaluation of Selected RO Membrane Using UF Permeate
Economic Assessment of Process
Economic Assessment of Process
Economic Assessment of Process

Evaluation of conventional emulsion breaking as a pretreatment for RO and membrane tests on this broken emulsion should be conducted to assess the appropriateness of this technology.

Economic assessments should be conducted using experimental data to evaluate the costs associated with each process.

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TECHNICAL AND ECONOMIC ASSESSMENT OF THE APPLICATION OF MEMBRANE TECHNOLOGY FOR TREATING OIL/WATER/MINERAL EMULSIONS

FINAL REPORT

VOLUME II - EXPERIMENTAL WORK AND ECONOMIC ASSESSMENTS

Report Prepared for

CANMET/ERL
Department of Energy, Mines, and Resources, CANADA
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ZENON ENVIRONMENTAL INC.

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SUMMARY

The prime objective of this study was to assess the technical and economic viability of applying membrane technology to the treatment of produced water emulsions.

A review of literature and present technology suggested that there is considerable potential for membrane application to produced water treatment. Ultrafiltration membranes have been used successfully in other industries for emulsified oil recovery and reverse osmosis systems have been successfully applied to the treatment of boiler feedwater.

Initial laboratory work involved a complete characterization of a sample of produced water emulsion. The concentration of oil in a very stable emulsion was 1700 mg/L. GC analysis showed the presence of 60 mg/L of naphthenic acid which is thought to contribute to the high stability of the emulsion.

Prescreening tests were conducted on twenty membranes manufactured from 6 different materials. One membrane of each type of material was chosen for further work based on the results of prescreening tests.

Membrane screening tests with the six best membranes were conducted at 25°C and 45°C with the pH of the feed adjusted to 5.5, left "as is" near 7.5, and adjusted to 9.5 and at 65°C with the pH left "as is". These tests showed clearly that flux rates increase at increasing temperatures and that higher pH appears to favour better flux rates. All membranes consistently removed virtually all oil from the feed. The TFC G10 membrane rejected the highest percentage of all

contaminants and maintained very stable flux rates during the tests. Membranes CHP Cat and PS E500 had generally higher flux rates than the TFC membrane but showed some decline of flux during the tests. Other membranes either had poor quality permeate or lower flux rates.

A batch concentration (to a 90% volume reduction) test and a long term stability test demonstrated that flux rates could be relatively high even at high volume reductions (0.62 m^3/m^2D at 80% volume reduction). The results also demonstrated that effective cleaning procedures will be critical to the economic viability of the application. Reverse osmosis tests demonstrated that RO membranes can in one simple step produce boiler feed water quality water from produced water which has had oil removed.

Economic assessments were conducted for UF/Sodium Zeolite softening, UF/RO, and induced gas floatation/filtration/RO processes for produced water treatment. The costs per thousand gallons treated, including capital amortization and operating costs were \$4.84, \$6.95, and \$7.06 respectively. The costs for tubular UF systems were prohibitive at the flux rates assessed in this study but may prove to be economical if flux rate improvements with tubular systems can be realized.

Further work should focus on assessing operation of configured membrane systems for potential flux rates and cleanability of such configured and larger scale systems. Higher temperature operation should also be explored as well as field testing to assess the impact of emulsion aging.

1.0 INTRODUCTION

This report by ZENON Environmental is submitted to CANMET division of the Department of Energy, Mines, and Resources, Canada in partial fulfillment of the requirements of the Contract, "Technical and Economic Assessment of the Application of Membrane Technology for Treating Oil/Water/Mineral Emulsions".

Volume I of this report contains a review of literature and industrial processes related to this topic. This volume (Volume II), contains the results of experimental work conducted by ZENON Environmental. Ultrafiltration (UF) membranes were evaluated in a small membrane test unit at different temperatures and pH. Selected membranes were tested in the small test unit at a ten fold concentration of the feed, over an extended period of time. Reverse osmosis membranes were also evaluated in a small test unit using permeate from the UF system and chemically demulsified produced water. This volume also contains the results of an economic assessment estimating the capital and operating costs for different processes treating produced water emulsions. These estimates are based on the results of experimental work, ZENON's in-house membrane system manufacturing experience, and contacts with suppliers of other process components.

The Appendices to this report are presented in Volume III.

2.0 EXPERIMENTAL PROCEDURES

2.1 Sample Characterization

2.1.1 Sample Description

Ten 45 gallon drums of wellhead emulsion were obtained from Texaco Canada's in-situ heavy oil recovery operation at Fort McMurry, Alberta. The drums each contained a layer of settled heavy oil sludge, a layer of light surface oil, and middle layer containing a stable oil in water emulsion. This oil in water emulsion represented approximately 90% by volume of the samples obtained. The surface oil layers from each of the drums were skimmed off and the emulsion layers were pumped into a large stainless steel mixing tank. This composite emulsion was mixed for several hours then transferred to clean 45 gallon drums for use throughout the experiments. Prior to samples being taken from a drum the contents were mixed to insure that a representative sample was being obtained. The composite emulsion, free of surface oil and settled sludge, was considered to be roughly equivalent to the effluent of a skim tank used in oil field operations. The removal of suspended oil in this manner is not difficult to carry out as a pretreatment for large-scale operation and can be expected to significantly improve membrane system operation.

In order to evaluate the variability of produced water three additional one liter samples of wellhead emulsion containing higher oil contents were also obtained from Texaco's Fort McMurry operation. These samples had been collected at the Fort McMurry pilot plant over a six month period in the latter half of 1983. The oil in water emulsions from these three samples were separated from the light surface oil layers and the settled heavy oils.

2.1.2 Particle Size Analysis

Physical characterization of the emulsion was conducted because of the importance of particle size in emulsion stability and membrane system operation.

Particle size distribution measurements were conducted by the Ontario Research Foundation on an unfiltered sample of the emulsion. A 1 mL portion of raw sample was also washed with 50 mL of toluene in a separatory funnel. To this, 9 mL of Isoton (a particle free aqueous electrolyte) was added to aid in particle dispersion and separation. Once the oil had been removed from the aqueous phase (as noted by a change in color of both the toluene and aqueous layers), the aqueous portion was collected for particle size analysis. Residual clay particles in the separatory funnel were washed through and collected with successive additions of Isoton.

Particle size measurement on both the raw sample and the toluene washed sample were conducted on a Coulter Counter TAII using a 70 micron aperture tube which provided an effective size range of 1.6 - 32+ micron. By preparing accurately diluted samples and metering a precise sample volume through the aperture tube, estimates of particle concentraton, within the range stated above, could be deduced. To arrive at the size distribution for the oil droplets, the toluene washed sample data (e.g. clay particles) were subtracted from the total or raw sample.

2.1.3 Emulsion Stability

The stability of the large volume sample of produced water emulsion was evaluated by examining the effect of pH, heating, freezing, and demulsifier addition on emulsion stability. The stability of the three higher oil content samples with respect to pH was also evaluated. These simple tests were conducted to provide background information on the technical feasibility of alternative processes for emulsion separation.

Effect of pH

In order to determine the stability of the emulsions with respect to pH, samples were pH adjusted to different values with NaOH or H_2SO_4.

The pH of 50 mL samples of the large volume of "typical" emulsion were adjusted to values of 2, 4, 4.5, 5.0, 5.5, 6, 8, 10, and 12. The samples were mixed and allowed to stand for several hours. The degree of separation over time and quality of supernatant were observed and recorded. Samples of the higher oil content emulsion had their pH adjusted to 3.0, 3.5 and 4.0. Again the effects on emulsion stability were observed.

Effect of Freezing

A 100 mL sample of emulsion was placed in a polyethelene container and put in a freezer for 24 hours. The container was removed from the freezer and allowed to warm to room temperature. The degree of separation and quality of supernatant obtained were observed.

Effect of Heating

An aliquot of emulsion was heated in a boiling water bath for a 1/2 hour period of time. The sample was then allowed to cool and the degree of separation and quality of supernatant obtained were observed.

Effects of Chemical Demulsifiers

Four samples of chemical demuslifiers were obtained from Witco Chemical. The four demulsifiers obtained were RTC 323, RTC 326, RTC 327, and RTC 330. General information on Witco's reverse demulsifiers and typical properties for RTC-323 are presented in Appendix A. These products are specified for breaking stable oil in water emulsions and are referred to by Witco as reverse demulsifiers. For the purpose of this report they will be referred to simply as demulsifiers.

Stock dilutions containing 10% by volume of each demulsifier were prepared with tap water and with a 2% ZnCl solution. One mL portions of each of the eight solutions were added to beakers containing 100 mL of produced water emulsion. The contents of the beakers were mixed and then allowed to stand. The degree of separation with time and quality of the supernatant were observed and recorded. The solutions which were successful in breaking the emulsion at a 10% concentration level were further diluted to 2.5% and 0.625% demulsifier in water or $ZnCl_2$ solution. The ability of these further diluted demulsifiers to break the emulsion was tested in the same manner as before.

2.1.4 Routine Chemical Analysis

The following analyses were conducted on the standard sample of emulsion at the start of experimental work and before the batch concentration tests:

pH,
Turbidity,
Conductivity,
Total Organic Carbon (TOC),
Filtered Organic Carbon (FOC),
Total Suspended Solids (TSS),
Total Suspended Oil (TSO),
Total Dissolved Solids (TDS),
Total Dissolved Oil (TDO),
Anions,
Cations,
Hardness (by calculation),
Dissolved Silica, and
Oil and Grease.

Initial analysis of the standard emulsion also included ethyl acetate extractable surfactant material as suggested by the scientific authority. The FOC, anion scan, cation scan, hardness, and silica measurements were done on samples which had been filtered prior to analysis. The procedures used for these analyses are presented in Appendix A.

These same analyses with the exception of oil and grease were performed on the three higher oil content emulsion samples.

2.1.5 GC and IR Analysis

Samples of the typical emulsion were prepared in three different ways for analysis by gas chromatography (GC) with a flame ionization detector (FID) and by infra-red spectrophotometry (IR). One hundred and fifty millilitres of the first sample was adjusted to pH 2 with concentrated hydrochloric acid and was then extracted with 50 mL of dichloromethane three times. The 150 mL of organic extract was then evaporated down to a volume of 10 mL and dried by passing it through a bed of Na_2SO_4. A portion of the sample was then injected into a Hewlett-Packard 5880A gas chromatograph with flame ionization detection. A separate portion of this sample was subjected to analysis with a Beckman infra-red spectrophotometer. Details of the GC and IR conditions are presented in Appendix B.

The pH of the second sample was adjusted to two and the supernatant was separated from the settled heavy oil phase. The aqueous phase of the sample was extracted with dichloromethane then subjected to the same GC and IR analysis as for the previous extracted sample.

In the third sample of emulsion the acid fraction was isolated prior to analysis. The pH of a sample of emulsion was adjusted to two and extracted with dichloromethane. This organic phase was then back extracted with aqueous sodium hydroxide. The aqueous base extract containing salts of organic acids was then reacidified and extracted with dichloromethane to yield the acid fraction in an organic phase. This organic phase was evaporated down and dried with Na_2SO_4 before GC and IR analyses.

Figure 2-1 shows a summary of the three methods of sample preparation used.

2.2 Membrane Prescreening Tests

2.2.1 Membrane Description and Criteria for Selection

A total of twenty membranes of different porosities and of six different materials were examined during the prescreening tests. Table 2-1 shows the membrane type, manufacturer, trade name, and molecular weight cut off or estimated pore size for each of the membranes tested.

All twenty membranes were evaluated for pure water and salt water flux, and 15 of the membranes were evaluated during the emulsion prescreening test. The polyethelene glycol (PEG) test was conducted using six membranes after emulsion prescreening tests had identifed the best membranes of each type. Membrane G10, CA-20K, E500, RC-51, Cat, and PVDF were tested during the PEG test.

The purpose of the prescreening tests was to identify the specific membrane of each type most appropriate for use in this application and to characterize the membranes with respect to the standard parameters of productivity, salt rejection, and molecular weight cut off.

The selection of the specific membranes of each type was based primarily on the results obtained from the emulsion prescreening test. The desired membrane characterisitics from the emulsion prescreening test were:

i) rejection of oil,

ii) a stable flux rate,

FIGURE 2 - 1

PREPARATION OF EMULSION SAMPLES FOR GC & IR ANALYSIS

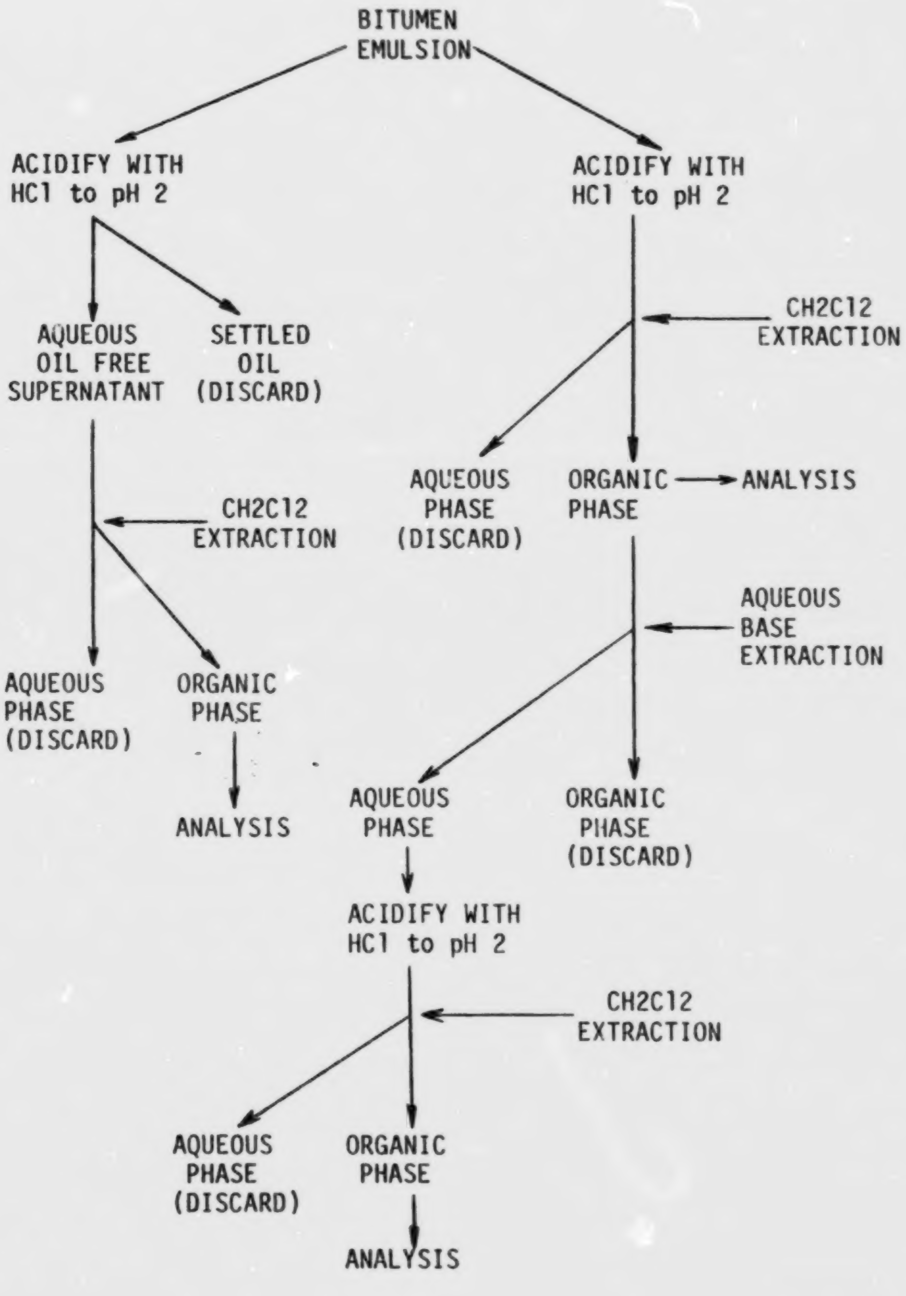

BITUMEN EMULSION
ACIDIFY WITH HCl to pH 2
ACIDIFY WITH HCl to pH 2
AQUEOUS OIL FREE SUPERNATANT
SETTLED OIL (DISCARD)
CH2Cl2 EXTRACTION
CH2Cl2 EXTRACTION
AQUEOUS PHASE (DISCARD)
ORGANIC PHASE
ANALYSIS
AQUEOUS PHASE (DISCARD)
ORGANIC PHASE
ANALYSIS
AQUEOUS BASE EXTRACTION
AQUEOUS PHASE
ORGANIC PHASE (DISCARD)
ACIDIFY WITH HCl to pH 2
CH2Cl2 EXTRACTION
AQUEOUS PHASE (DISCARD)
ORGANIC PHASE
ANALYSIS

TABLE 2 - 1

MANUFACTURERS DESCRIPTION OF MEMBRANES USED

MEMBRANE TYPE	MANUFACTURER	TRADE NAME	MOLECULAR WT CUT OFF OR PORE SIZE
CELLULOSE ACETATE	OSMONICS	SEPA-0(CA)	1000 MW
	OSMONICS	SEPA-20(CA)	20000 MW
	OSMONICS	SEPA-50(CA)	50000 MW
	DESALINATION	CA - UF	20000 MW
POLYSULFONE	OSMONICS	SEPA-0K(PS)	1000 MW
	OSMONICS	SEPA-20K(PS)	20000 MW
	OSMONICS	SEPA-50K(PS)	50000 MW
	DESALINATION	E500	500 Angstroms
CHP (proprietary)	ZENON	Cat	20000 MW
		B2	10000 MW
		B3	10000 MW
CELLULOSE	S. AND S.	RC - 51	50 Angstroms
	*	12	NA
	*	13	NA
	*	23	NA
THIN FILM COMPOSITE	DESALINATION	G5	5 Angstroms
	DESALINATION	G10	10 Angstroms
	DESALINATION	G20	20 Angstroms
	DESALINATION	G50	50 Angstroms
POLYVINYLDIFLUORIDE	ABCOR	HF - 100 (P521)	10000 MW

S. AND S. (SCHLEICHER AND SCHUELL)
* SUPPLIED BY DR. T MATSUURA , NRC DIV. OF CHEM.

iii) a high flux rate, and

iv) rejection of organic carbon.

2.2.2 Pure Water Flux Test

2.2.2.1 Apparatus

The pure water flux tests were conducted using three National Research Council (NRC) test cells each with a membrane surface area of 13.3 cm^2. The flow in these cells enters from the outer edge of a circular membrane sheet and flows across the membrane surface to an exit port at the centre. The membrane is supported on a sintered stainless steel support plate. Figure 2-2 shows the design of the NRC tests cell units.

The test system had a 20 L process tank which fed an Osmonics, Tonka Flo centrifuged pump model #SB520X4200. The flow to each of the cells was maintained at 2 L/min. by allowing excess flow from the pump to return to the process tank via a by-pass line. The flow in the by-pass line was regulated manually with a needle valve. The pressure in the cells was maintained at the desired level using a back pressure regulating valve downstream from the cells. The temperature of the water in the system was controlled using cooling coils in the process tank. Heat was provided by the process pump. Figure 2-3 shows a flow schematic for the test apparatus.

2.2.2.2 Procedure

Samples of the membranes were placed in the NRC cells and the system was operated using pure water. The pure water was reverse osmosis (RO) permeate generated using tap water as a feed.

Figure 2 - 2

NRC TEST CELLS

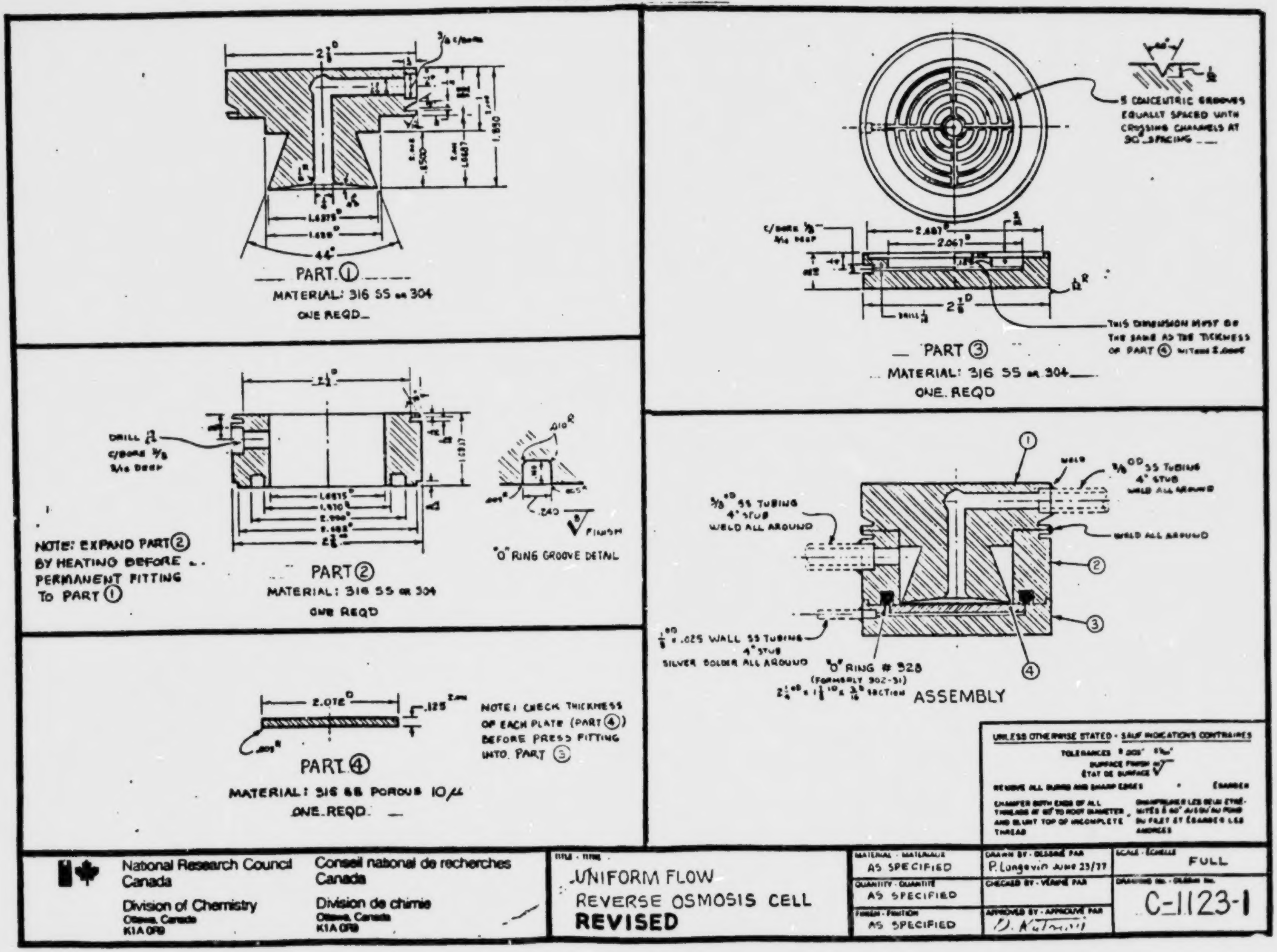
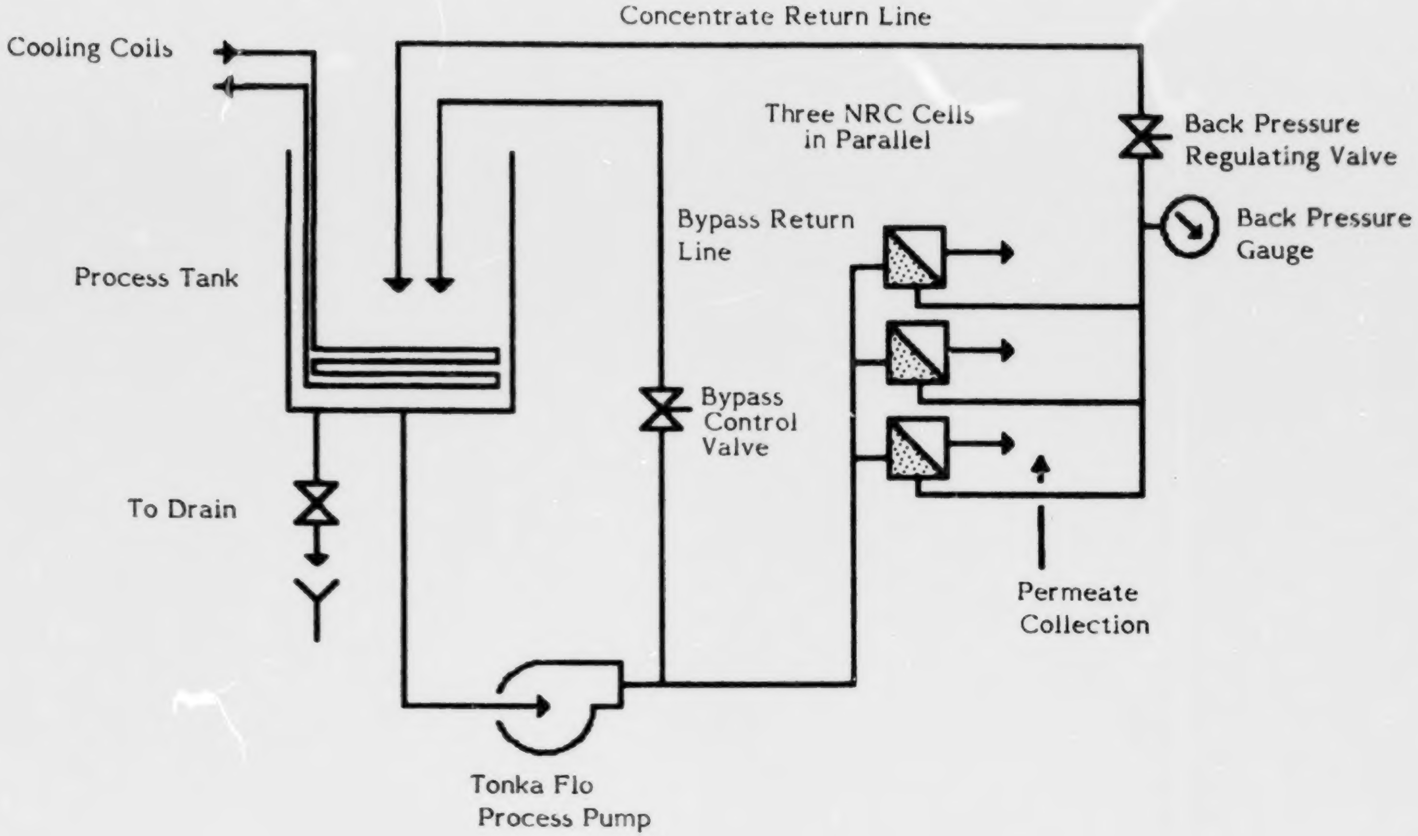

PART ①
MATERIAL: 316 SS or 304
ONE REQD
$2\frac{1}{8}$ D
3/8 C/BORE
1.850
.8500
1.6687
1.6875 D
1.480 D
44°
PART ②
MATERIAL: 316 SS or 304
ONE REQ'D
DRILL $\frac{1}{16}$ C/BORE 3/8 3/16 DEEP
$2\frac{1}{8}$ D
1.6875
1.910
2.000
2.482
$2\frac{3}{8}$
NOTE: EXPAND PART ② BY HEATING BEFORE PERMANENT FITTING TO PART ①
.010 R
.240
FINISH
"O" RING GROOVE DETAIL
PART ③
MATERIAL: 316 SS or 304
ONE REQD
5 CONCENTRIC GROOVES EQUALLY SPACED WITH CROSSING CHANNELS AT 30° SPACING
C/BORE 3/8 3/16 DEEP
2.487 D
2.067 D
$2\frac{3}{8}$ D
DRILL $\frac{1}{16}$
THIS DIMENSION MUST BE THE SAME AS THE THICKNESS OF PART ④ WITHIN ±.0005
PART ④
MATERIAL: 316 SS POROUS 10μ
ONE REQD
2.072 D
.125
NOTE: CHECK THICKNESS OF EACH PLATE (PART ④) BEFORE PRESS FITTING INTO PART ③
ASSEMBLY
①
②
③
④
WELD
3/8 OD SS TUBING 4" STUB WELD ALL AROUND
WELD ALL AROUND
3/8 OD SS TUBING 4" STUB WELD ALL AROUND
1/8 OD × .025 WALL SS TUBING 4" STUB SILVER SOLDER ALL AROUND
"O" RING # 328 (FORMERLY 302-31) $2\frac{1}{4}$ ID × $1\frac{1}{8}$ OD × $\frac{3}{16}$ SECTION
UNLESS OTHERWISE STATED · SAUF INDICATIONS CONTRAIRES
TOLERANCES ±.005" ±1/64"
SURFACE FINISH · ÉTAT DE SURFACE
REMOVE ALL BURRS AND SHARP EDGES · ÉBARBER
CHAMFER BOTH ENDS OF ALL THREADS AT 45° TO ROOT DIAMETER AND BLUNT TOP OF INCOMPLETE THREAD · CHANFREINER LES DEUX EXTRÉMITÉS À 45° JUSQU'AU FOND DU FILET ET ÉBARBER LES AMORCES
National Research Council Canada
Conseil national de recherches Canada
Division of Chemistry Ottawa, Canada K1A 0R9
Division de chimie Ottawa, Canada K1A 0R9
TITLE · TITRE
UNIFORM FLOW REVERSE OSMOSIS CELL REVISED
MATERIAL · MATÉRIAUX AS SPECIFIED
QUANTITY · QUANTITÉ AS SPECIFIED
FINISH · FINITION AS SPECIFIED
DRAWN BY · DESSINÉ PAR P. Longevin June 23/77
CHECKED BY · VÉRIFIÉ PAR
APPROVED BY · APPROUVÉ PAR D. Kutowy
SCALE · ÉCHELLE FULL
DRAWING No. · DESSIN No. C-1123-1

Figure 2 - 3

FLOW SCHEMATIC FOR PURE WATER TEST APPARATUS


Concentrate Return Line
Cooling Coils
Three NRC Cells
in Parallel
Back Pressure
Regulating Valve
Back Pressure
Gauge
Bypass Return
Line
Process Tank
Bypass
Control
Valve
To Drain
Permeate
Collection
Tonka Flo
Process Pump

The system was operated with a pressure of 100 psi and at a temperature of 25°C for three hours. The system was allowed to operate for one hour before measurements of the flux rate were taken. This time allowed the flux rates to stabilize after initial compaction of the membrane. Flux rates were monitored and recorded periodically over the remainder of the test period.

2.2.3 Salt Water Flux Test

2.2.3.1 Apparatus

The apparatus used for the salt water flux test was the same as was used for the Pure Water Flux Tests described in section 2.2.2.1.

2.2.3.2 Procedure

Samples of the membranes were placed in the NRC test cells and the system was operated using a 1500 mg/L solution of NaCl as a feed to the cells. Ten liters of the salt solution was prepared using RO permeate with the appropriate amount of NaCl. The system was operated with a pressure of 100 psi and at a temperature of 25°C for three hours.

The conductivity of the feed and permeate and the flux rates were measured at regular intervals during the test. The procedure for conductivity measurement was the same as outlined for emulsion characterization in Appendix A.

The percent rejection of NaCl is calculated from:

$$R = \frac{C_f - C_p}{Cf} \times 100 \qquad (2\text{-}1)$$

where R = percent rejection

C_f = conductivity of the feed

C_p = conductivity of the permeate

The concentration of salt was maintained constant throughout the test by recycling both the permeate and the concentrate back to the process tank.

2.2.4 Emulsion Prescreening Tests

2.2.4.1 Apparatus

The emulsion prescreening tests were conducted using six NRC test cells arranged in parallel.

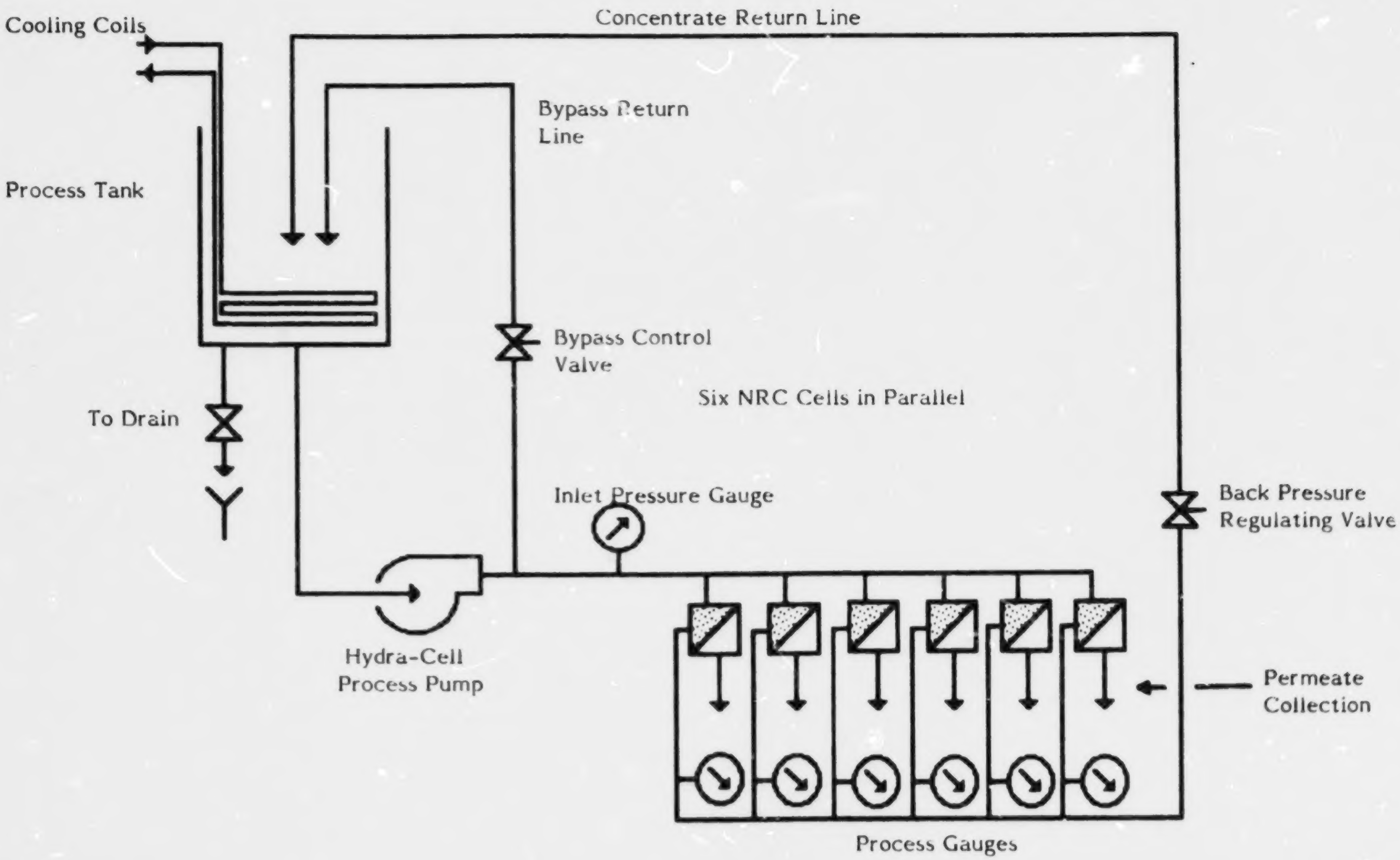
The 20 L process tank fed a hydra-cell pump model D-105516. The pump capacity was rated at 10 USGPM at 100 psi outlet pressure. As before, the flow to each of the cells was maintained at 2 L/min by manually adjusting a needle valve on a by-pass line from the pump outlet to the process tank. The temperature was controlled by manually adjusting the flow of water in cooling coils in the process tank. The pressure was controlled manually with a back pressure regulating valve on the concentrate return line. The pressure was measured on both the inlet and outlet to the NRC modules. A flow schematic for this system is shown in Figure 2-4.

2.2.4.2 Procedure

Samples of the membranes were placed in the NRC test cells and the system was operated using bitumen emulsion as a feed.

Figure 2 - 4

FLOW SCHEMATIC FOR EMULSION STABILITY TEST APPARATUS


Cooling Coils
Concentrate Return Line
Bypass Return Line
Process Tank
Bypass Control Valve
Six NRC Cells in Parallel
To Drain
Inlet Pressure Gauge
Back Pressure Regulating Valve
Hydra-Cell Process Pump
Permeate Collection
Process Gauges

The temperature of the feed was maintained at 25°C and the pressure to the cells was 100 psi. The tests were conducted for a period of 3 1/2 hours. Flux rates and samples of cell permeates were taken every 1/2 hour. The samples were analyzed for TOC using the procedure outlined in Appendix A. At the completion of the test samples of the feed and composite samples of permeate from each cell were analyzed for pH, conductivity, and TDS as per the procedures described in Appendix A.

2.2.5 Polyethelene Glycol (PEG) Test

2.2.5.1 Apparatus

The polyethelene glycol (PEG) tests were conducted using 6 NRC test cells set up in series.

The test system had a 20 L process tank which fed a Moyno Pump model ZM4 type SSF with a viton stator. As with previous systems the flow to the cells was maintained at 2 L/min by manually adjusting a needle valve on a by-pass line from the pump outlet to the process tank. The temperature was controlled by manually adjusting the flow of water in cooling coils in the process tank. The pressure to the cells was controlled with a back pressure regulating valve downstream of the final cell in series. The pressure at the outlet from the final cell was maintained at 100 psi and the pressure at the inlet of the first cell generally remained near 150 psi. A flow schematic for the system is presented in Figure 2-5.

2.2.5.2 Procedure

Samples of the membranes were placed in the NRC test cells and the system was operated using a 500 mg/L solution of a 6000 MW polyethelene glycol. The temperature of the system was maintained at 25°C and the pressure at the outlet was 100 psi. Tests were conducted for a period of one hour, during which time flux rates and samples of the permeate were taken at regular intervals. The samples were analyzed for TOC as before.

The percent rejection of PEG for each membrane was determined using.

$$R = \frac{C_f - C_p}{C_f} \times 100 \qquad (2\text{-}2)$$

where R = Percent Rejection

C_f = TOC feed concentration

C_p = TOC permeate concentration

The molecular weight cut off values for the membranes were determined by a comparison of percent rejections obtained with this test and those obtained with membranes of known molecular weight cut offs during previous membrane testing by ZENON staff.

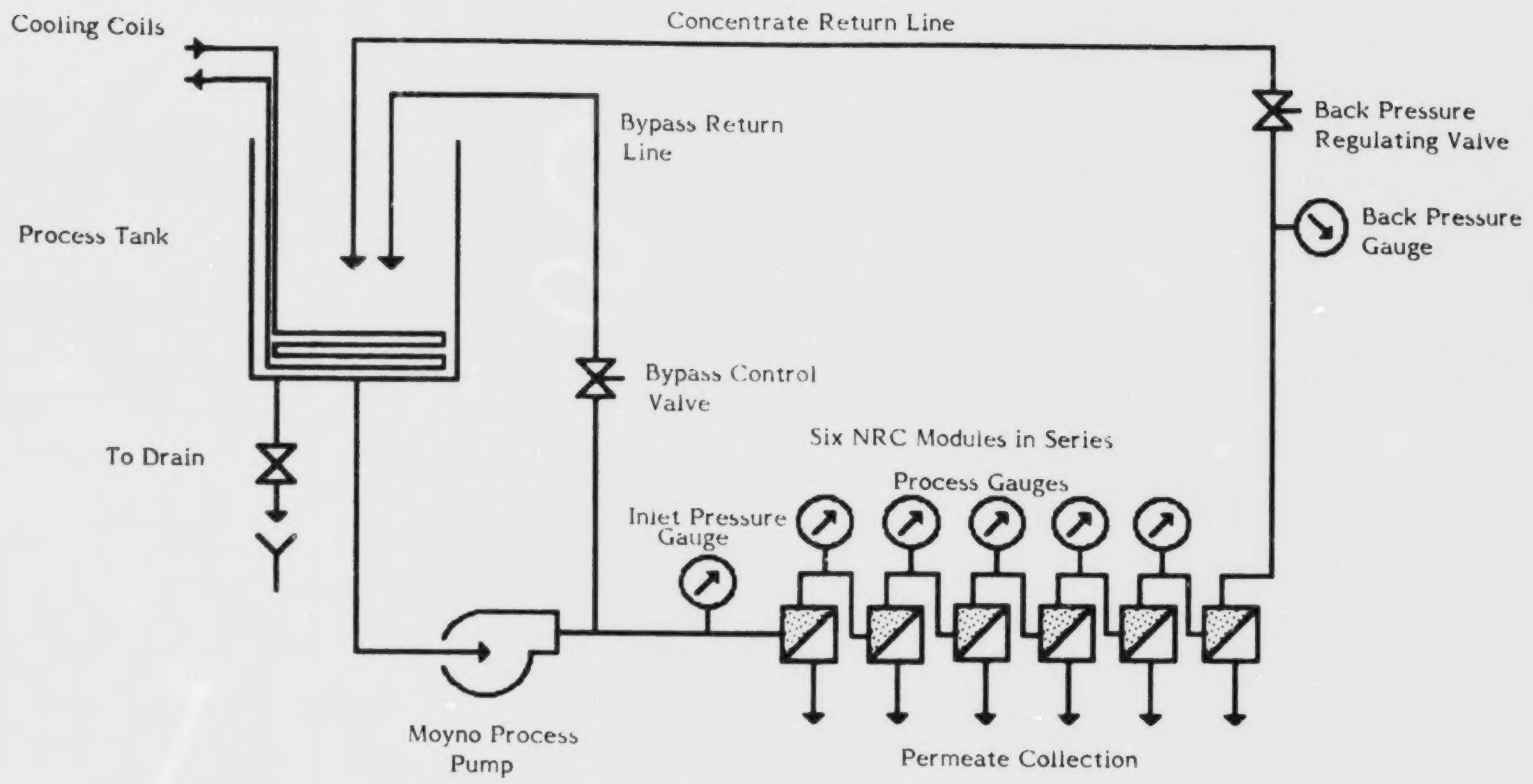
2.3 Membrane Screening Tests

2.3.1 Apparatus

The test apparatus used for the membrane screening tests was the same as that used for the PEG tests described in Section 2.2.5.1. A flow schematic for the system is shown in Figure 2-5.

Figure 2 - 5

FLOW SCHEMATIC FOR POLYETHELENE GLYCOL TEST APPARATUS


Cooling Coils
Concentrate Return Line
Bypass Return Line
Back Pressure Regulating Valve
Back Pressure Gauge
Process Tank
Bypass Control Valve
To Drain
Six NRC Modules in Series
Process Gauges
Inlet Pressure Gauge
Moyno Process Pump
Permeate Collection

The Moyno pump was substituted for the Hydra-Cell for these tests because the values on the Hydra-Cell pump plugged with bitumen when operated for more than three or four hours. The particular Moyno used was limited to operation below 70°C because differential expansion of the components placed excessive strain on the pump motor. Moyno pumps capable of operating at higher temperatures are available.

2.3.2 Procedure

A total of 48 screening tests were conducted using the six different membranes chosen in the prescreening tests. In order to examine the effect of temperature and pH on membrane flux rates and permeate quality the feed to the system was maintained at different temperatures and adjusted to different pH values.

The temperature of the emulsion feed was controlled at 25°C, 45°C, or 65°C by adjusting the flow of cooling water to the coils in the process tank. Sixty five °C is considered to be the upper limit of most membrane system operation. The pH of the emulsion was either adjusted downwards to 5.5 with H_2SO_4, left as is at about 7.5, or adjusted upwards to 9.5 with NaOH. Table 2-2 shows a summary of all the screening tests performed and the operating conditions for each test.

Each of the screening tests was run for a period of 24 hours. During that time the flux rates were monitored every hour except for a period of about eight hours during the night. Composite samples of the permeate from each membrane and samples of the initial and final feed were collected and analyzed for the following parameters:

TABLE 2 - 2

OPERATING CONDITIONS FOR MEMBRANE SCREENING TESTS

RUN NO	TEMPERATURE (C)	pH	MEMBRANES
1	25	AS RECEIVED	TFC,CHP,PS,CA,PVDF,C
2	45	AS RECEIVED	TFC,CHP,PS,CA,PVDF,C
3	25	5.5	TFC,CHP,PS,CA,PVDF,C
4	45	5.5	TFC,CHP,PS,CA,PVDF,C
5	25	9.5	TFC,CHP,PS,CA,PVDF,C
6	45	9.5	TFC,CHP,PS,CA,PVDF,C
7	65	AS RECEIVED	TFC,CHP,PS,CA,PVDF,CHP
8	65	AS RECEIVED	PS ALL CELLS

TFC	THIN FILM COMPOSITE
CHP	ZENON MEMBRANE
PS	POLYSULPHONE
CA	CELLULOSE ACETATE
PVDF	POLYVINYL DIFLUORIDE
C	CELLULOSE

and final feed were collected and analyzed for the following parameters:

- pH,
- Conductivity,
- Turbidity,
- TOC,
- TDS,
- Ca and Mg,
- Hardness, and
- Dissolved Silica.

In addition to the regular analysis, the feed samples were analyzed for total suspended solids and a composite sample of the permeate from run number eight with PS membranes in all cells was subjected to detailed analysis for organics by GC & IR. The procedures for these analyses are presented in Appendix A.

Since each of the six cells in series had a slightly different applied pressure at the membrane surface the flux rates were adjusted to be equivalent to the flux rates at the inlet pressure using the formula

$$Q_i = Q_o \left(1 + \frac{P_i - P_o}{P_i}\right) \qquad (2\text{-}3)$$

where Q_i = Corrected Flux Rates or Flux Rate at Inlet Pressure

Q_o = Observed Flux Rate

P_i = Inlet Pressure

P_o = Observed Pressure in Cell

2.4 Batch Concentration Test

2.4.1 Apparatus

The apparatus for the batch concentration test was similar to that used for the PEG and membrane screening tests except that the by-pass line was set up to allow the by-pass flow to return to the process tank as before or to flow through a large plate and frame flat cell membrane test unit. This large cell unit contained four membranes each with a membrane surface area of 177 cm^2. Within the unit the four membranes were set up in parallel flow. The flow across each membrane entered one side of the rectangular membrane surface and flowed across the membrane and out near the side of the rectangle opposite to the feed side. This test unit was designed and manufactured by ZENON Environmental. Figure 2-6 shows the construction of the large cell test unit. Valves were put in place so that the by-pass flow could be directed to return directly to the process tank or to the process tank after passing through the large cell test unit. This large flat cell unit was used in order to achieve a higher concentration of the original feed more quickly than would have been possible using the NRC cells alone. A flow schematic for this equipment is shown in Figure 2-7.

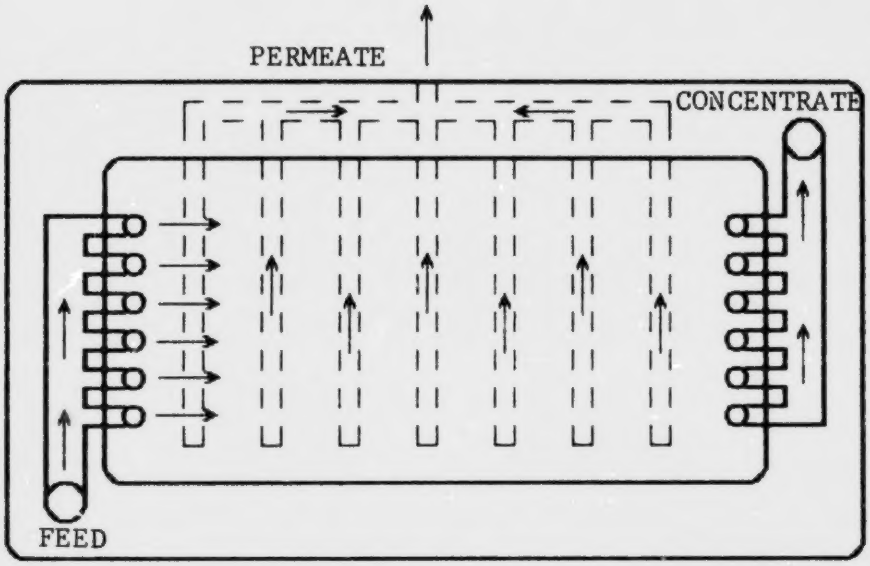
2.4.2 Procedure

Membranes G10, E500, and Cat were each placed in two NRC test cells in the test apparatus. Sheets of PVDF were also placed in the large cell test unit.

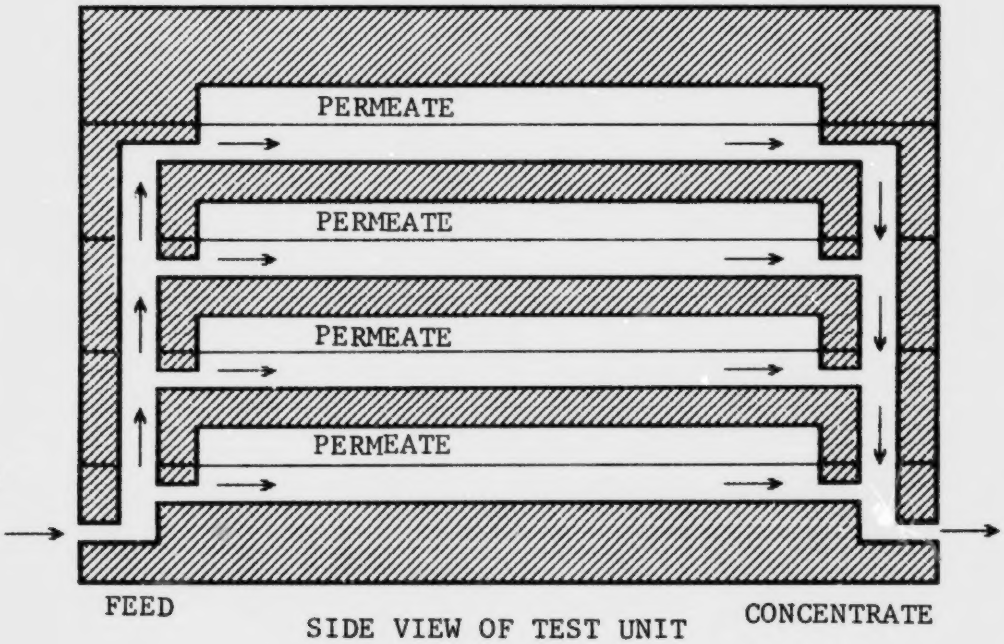
The pH of the bitumen emulsion feed was not adjusted and the temperature was maintained at 65°C. A flow to the NRC test cells of 2

FIGURE 2 - 6

PLATE AND FRAME MEMBRANE TEST UNIT


PERMEATE
CONCENTRATE
FEED

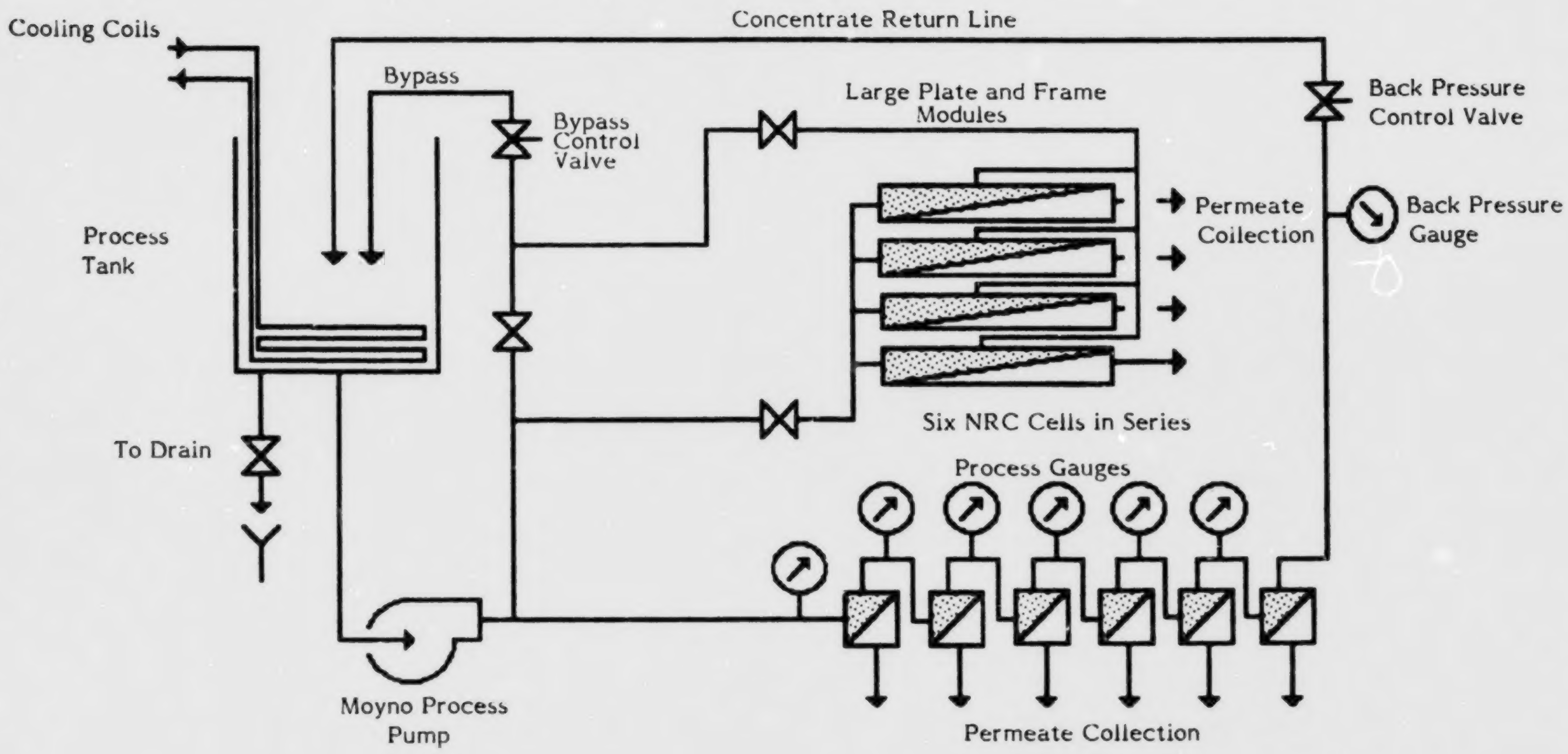
TOP VIEW OF INTERMEDIATE PLATE


PERMEATE
PERMEATE
PERMEATE
PERMEATE
FEED
CONCENTRATE

SIDE VIEW OF TEST UNIT

FIGURE 2 - 7

FLOW SCHEMATIC FOR BATCH CONCENTRATION TESTS


Cooling Coils
Concentrate Return Line
Bypass
Bypass Control Valve
Large Plate and Frame Modules
Back Pressure Control Valve
Process Tank
Permeate Collection
Back Pressure Gauge
Six NRC Cells in Series
To Drain
Process Gauges
Moyno Process Pump
Permeate Collection

L/min was maintained and the pressure at the outlet of the last cell was 100 psi.

The system was operated for a period of 50 hours with the permeate withdrawn from the system and the concentrate recycled to the process tank until the desired 90% volume reduction was obtained. During that time flux rates were monitored and samples of the permeate from the NRC test cells were taken at the start of the run, at 30%, 60%, and 90% volume reduction levels. These samples were analyzed for turbidity and TOC using the procedure outlined in Appendix A. Samples of the concentrated emulsion were removed from the process tank and allowed to stand for a period of several hours to determine if emulsion breaking was occurring.

The flux rates obtained for the NRC cells were adjusted for pressure by the same procedure used for the screening tests outlined in equation 2-3.

2.5 Long Term Stability Test

2.5.1 Apparatus

The apparatus for the long term stability test was the same as was used for the batch concentration test. During the long term stability test the valves on the by-pass line were set so that all the flow in the by-pass line returned directly to the process tank and did not pass through the large flat test cells.

2.5.2 Procedure

Once a 90% volume reduction was achieved in the batch concentration test the long term stability test began. The membranes and operating conditions remained the same as during the concentration test. The flow in the system was changed so that the by-pass flow returned directly to the process tank. Both the permeate and the concentrate were also recycled to the process tank.

The system was operated for 300 hours after the 90% volume reduction was achieved. During this time the flux rates were monitored and samples of permeate from each of the membranes and from the feed tank were taken after one week and at the end of two weeks of operation. These samples were analyzed using procedures outlined in Appendix A for the following:

pH,
Conductivity,
Turbidity,
TOC,
TDS,
Dissolved Silica,
Ca and Mg, and
Hardness,

The samples of feed were also analyzed for total suspended solids (TSS) and total suspended oil (TSO). The flux rates obtained were adjusted for pressure using equation 2-3.

2.6 Reverse Osmosis Tests

2.6.1 Apparatus for RO Tests

The reverse osmosis tests were all conducted using the large plate and frame flat cell membrane test unit which was used for the batch concentration test described in Section 2.4.1. Figure 2-6 shows the construction of this membrane test unit.

The test system had a 20 L process tank equipped with cooling coils for maintaining the temperature of the feed at the desired level. The flow to the cooling coils was controlled automatically by a solenoid valve connected to a thermocouple in the process tank. The process tank fed a Hydra Cell Pump and the flow to the test unit was maintained at approximately 15 L/min by manually adjusting a needle valve on the by-pass line from the pump outlet to the process tank. The pressure was controlled manually with a back pressure regulating valve on the concentrate outlet of the test unit. The pressure at the outlet of the test unit was monitored.

A flow schematic for this system is shown in Figure 2-8.

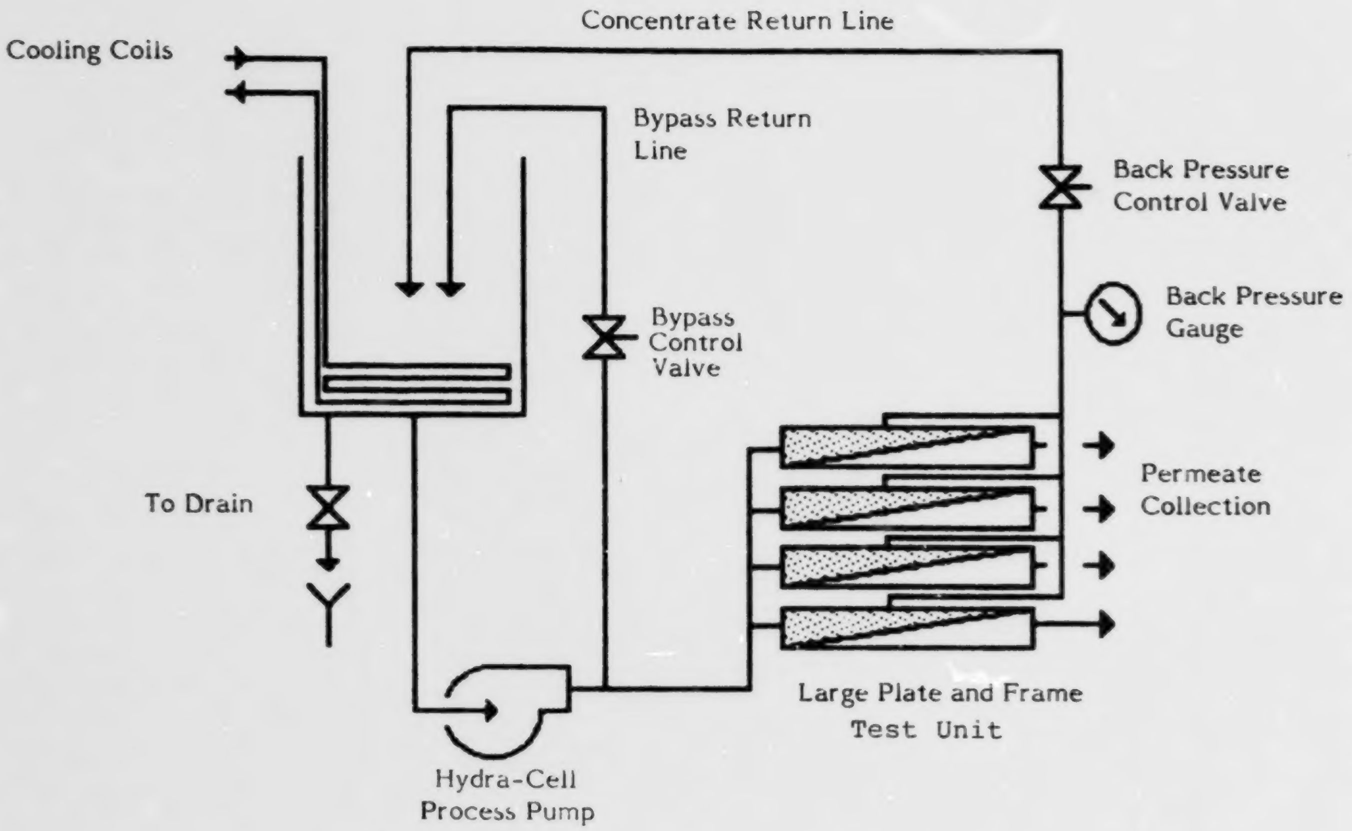
2.6.2 Reverse Osmosis Screening Test

Two different reverse osmosis membranes, a Filmtec FT30, thin film composite membrane and a cellulose acetate membrane (desalination), were used during the RO screening test. Sheets of each membrane were placed in two of the cells in the test unit.

A salt water flux test was performed on the cells in the test unit before testing with the UF permeate. A 1500 mg/L solution NaCl in RO water was feed to the system for a 4 1/2 hour period. The temperature of the feed was maintained at 25°C and the pressure to the cells was 400 psi. The flux rates and conductivity of the permeates from each cell and of the feed water were measured every half hour. The salt rejections were calculated using equation 2-1.

FIGURE 2 - 8

FLOW SCHEMATIC FOR R.O. TESTS


Concentrate Return Line
Cooling Coils
Bypass Return Line
Back Pressure Control Valve
Back Pressure Gauge
Bypass Control Valve
To Drain
Permeate Collection
Large Plate and Frame Test Unit
Hydra-Cell Process Pump

For the RO screening test the feed to the system was a composite sample of permeate generated during the final UF membrane screening test using a polysulfone membrane in each of the cells. The test was run for 4 1/2 hours with the temperature and pressure maintained at 25°C and 400 psi respectively. The conductivity and pH of the permeate from each cell and of the feed were measured every half hour. Flux rates were also measured every half hour.

A salt water flux test was performed on the membranes after completion of the test with UF permeate. The procedure used was the same as was used for the initial salt water test.

2.6.3 Reverse Osmosis Test on UF Permeate

Ultrafiltration permeate was generated by operating the large flat cell test unit with four sheets of the G10-TFC membrane. The operating temperature was 65°C and the pH of the emulsion feed was not adjusted. The system was operated until sufficient permeate was available for the RO test.

For the RO test, sheets of the Filmtec FT30, thin film composite seawater membrane were placed in each of the four cells of the test unit. An initial salt water flux test was performed on the test unit using the same procedure as was used during the screening test.

During the RO test on UF permeate the temperature of the feed was maintained at 25°C and the pressure to the cells was 400 psi. The test was run for just over five hours until a 90% volume reduction had been obtained. During the run flux rates were taken every 15 minutes and the conductivity of the permeate from each cell and of the feed were also measured every 15 minutes. The pH of the permeates and feed were measured every 1/2 hour.

Samples of the feed were taken before the tests began and at the completion of the run. Samples of the combined permeate from the four cells were taken after 30 min and when 30%, 60%, and 90% volume reductions had been achieved. All six of these samples were analyzed for the following:

pH,
Conductivity,
Turbidity,
TOC,
TDS,
Dissolved Silica,
Anions
Cations, and
Hardness

Samples of the final feed and the permeate at a 90% volume reduction were also analyzed for alkalinity.

A salt water flux test was also performed on the membrane system after completion of the test with UF permeate. The procedure used was the same as that used previously.

2.6.4 Reverse Osmosis Test on Demulsified Bitumen Emulsion

Demulsified bitumen emulsion was obtained by treating a 40 L volume of emulsion with 250 ppm of the chemical demulsifier RTC 323. The emulsion was mixed and the heavy oil was allowed to settle.

The supernatant was then drained off and passed through a five micron cartridge filter.

The remainder of the procedure for the reverse osmosis test on the demulsified bitumen emulsion was identical to that used for the RO test on UF permeate.

3.0 RESULTS AND DISCUSSION

3.1 Sample Characterization

3.1.1 Particle Size Analysis

The particle size analysis on the raw emulsion yielded a mean particle size of 7.8 microns based on particle volume and 2.8 microns based on population. The clay anaysis was conducted on material which could not be extracted with toluene. The mean particle size of this material was 9.8 microns based on particle volume and 2.5 microns based on population. The clay material made up only 7% of the total particle population of raw emulsion. Particle size data for other sources of produced water emulsion are not available for comparison but the very fine droplet size is indicative of very stable emulsions. Particles in the 7-10 micron range will be rejected easily by a membrane with pore sizes less than 0.5 microns (500 angstroms).

Table 3-1 shows a summary of the results of the particle size analysis on the bitumen emulsion. The complete results of the analysis are presented in Appendix B.

3.1.2 Emulsion Stability

Effect of pH

Samples of the standard bitumen emulsion showed no sign of emulsion destabilization at pH values of 4.5, 5.0, 5.5, 6, 8, 10, or 12. When the pH of a sample was adjusted to 4.0 the emulsion broke and a heavy oil layer settled within 2 hours. The supernatant was relatively free of suspended particles but retained a yellow tint. The settled oil compacted to about 10% of the total volume of sample. Another sample of emulsion in which pH was adjusted to 2 also broke

TABLE 3 - 1

RESULTS OF PARTICLE SIZE ANALYSIS
OF PRODUCED WATER EMULSION

FRACTION	PARTICLES PER (ml) OF SOLUTION	MEAN PARTICLE SIZE (MICRONS) BASED ON VOL %	MEAN PARTICLE SIZE (MICRONS) BASED ON POPULATION
RAW	4,744,000	7.8	2.8
CLAY (TOLUENE WASHED)	341,000	9.8	2.5
OIL (RAW-CLAY)	4,403,000	7.8	2.5

and a heavy oil layer settled within one hour. The supernatant from the pH 2 sample was clearer than that of the pH 4 sample and had a less intense yellow tint.

Samples of the three high oil content bitumen emulsions (refered to as P35, P33, and PI-11) each had their pH adjusted to 3.0, 3.5, and 4.0. Sample P35 showed no signs of separation at pH 4 or pH 3.5 and only slight separation at a pH of 3. Sample P33 showed no separation at pH 4, slight separation at pH 3.5, and clear separation at pH 3.0. Sample PI-11 showed no separation at pH 4 or pH 3.5 but clear separation at pH 3.0.

Table 3-2 shows a summary of the emulsion stability tests with respect to pH. It is beleived that the emulsions break at low pH as the polar surfactant which is substantially responsible for maintaining the stable emulsion becomes inactive at low pH. It is certainly technically feasible to break the emulsion entirely by lowering the pH substantially but this practice may be undesirable for downstream processes. Slight pH adjustment (± 2 pH units) to help destabilize emulsions is often employed in industrial oily water separations using membranes. The combination of a destabilized and concentrated emulsion often allows for oil to be skimmed off in the process tank. Such an approach was not useful in this case because a more drastic pH reduction was needed to effect any change in emulsion stability.

TABLE 3 - 2

STABILITY OF BITUMEN EMULSIONS WITH RESPECT TO pH

pH	STANDARD EMULSION	HIGH OIL P35	HIGH OIL P33	HIGH OIL PI-11
4.5 AND ABOVE	NO SEPARATION	NOT DONE	NOT DONE	NOT DONE
4.0	CLEAR SEPARATION	NO SEPARATION	NO SEPARATION	NO SEPARATION
3.5	NOT DONE	NO SEPARATION	SLIGHT SEPARATION	NO SEPARATION
3.0	NOT DONE	SLIGHT SEPARATION	CLEAR SEPARATION	CLEAR SEPARATION
2.0	CLEAR SEPARATION	NOT DONE	NOT DONE	NOT DONE

Effect of Freezing

The sample of emulsion which was frozen and allowed to thaw had a distinct heavy oil layer which settled out but the upper layer was still dark brown and somewhat cloudy.

Effect of Heating

The sample of emulsion which was heated to 100°C for a half hour showed little sign of emulsion destabilization. Increasing emulsion temperature will frequently destabilize less tightly bound emulsions.

Effect of Chemical Demulsifiers

Chemical demulsifiers are used commercially in many applications including produced water treatment. The concentration of demulsifiers required is critical to the economics of large systems. The chemical demulsifier RTC 323 was successful in breaking the emulsion at concentration levels of 1000 and 250 ppm. Clear separation of the heavy oil and the supernatant occured within 10 minutes of the addition of the demulsifier. At a concentration level of 62.5 ppm, the demulsifier had no effect on the emulsion even after several hours. Demulsifiers RTC 326, RTC 327, and RTC 330 each had no effect on the emulsion at concentration levels of 1000 ppm.

It had been suggested by the supplier that the emulsion breaking capabilities of Witco demulsifiers were enhanced by the presence of $ZnCl_2$ in solution.

Chemical demulsifiers complexed with $ZnCl_2$ acted differently from uncomplexed demulsifiers. The emulsion breaking mechanism of RTC 323 was somehow blocked by the $ZnCl_2$ in solution. RTC 323 in solution with $ZnCl_2$ was ineffective in breaking the bitumen emulsion.

RTC 326 and RTC 330 required the $ZnCl_2$ to function as demulsifiers. RTC 326 at 1000 ppm with $ZnCl_2$ at 200 ppm broke the bitumen emulsion and clear separation of the heavy oil occured within about 15 minutes. RTC 326 at 250 ppm with $ZnCl_2$ at 50 ppm was also effective in breaking the emulsion. At levels of 62.5 and 12.5 ppm no emulsion breaking occured. RTC 330 acted in an identical manner to RTC 326 and worked well at 1000 ppm with 200 ppm of $ZnCl_2$ and at 250 ppm with 50 ppm of $ZnCl_2$ but did nothing at 62.5 ppm with 12.5 ppm $ZnCl_2$. RTC 327 had no effect on emulsion stability at 1000 ppm complexed with 200 ppm $ZnCl_2$. Table 3-3 shows a summary of the results of demulsifier tests on the bitumen emulsion.

Since RTC 323 alone performed as well as RTC 326 and RTC 330 complexed with $ZnCl_2$, RTC 323 was chosen as the demulsifier to be used to break the emulsion for the reverse osmosis test on the broken emulsion.

3.1.3 Standard Analytical Parameters

Results of the analyses of bitumen emulsions are presented in Table 3-4. The table includes results of analyses performed on the standard emulsion at the start of experimental work and at the start of the batch concentration tests and that performed on the three high oil content samples. Also presented in the table are typical values for such emulsions as given by CH2M HILL (1982).

TABLE 3 - 3

EFFECT OF CHEMICAL DEMULSIFIERS ON BITUMEN EMULSION

DEMULSIFIER	CONCENTRATION	RESULT
RTC 323	1000 ppm	Clear Separation, oil sinks
RTC 323	250 ppm	Clear Separation, oil sinks
RTC 323	62.5 ppm	No separation
RTC 326	1000 ppm	No separation
RTC 327	1000 ppm	No separation
RTC 330	1000 ppm	No separation
RTC 323 +ZnCl 2	1000 ppm 200 ppm	No separation
RTC 326 +ZnCl 2	1000 ppm 200 ppm	Clear Separation, oil sinks
RTC 326 +ZnCl 2	250 ppm 50 ppm	Clear Separation, oil sinks
RTC 326 +ZnCl 2	62.5 ppm 12 ppm	No separation
RTC 327 +ZnCl 2	1000 ppm 200 ppm	No separation
RTC 330 +ZnCl 2	1000 ppm 200 ppm	Clear Separation, oil sinks
RTC 330 +ZnCl 2	250 ppm 50 ppm	Clear Separation, oil sinks
RTC 330 +ZnCl 2	62.5 ppm 12 ppm	No separation

Witco Chemical refers to agents which are designed for separating oil in water emulsions as reverse demulsifiers. The demulsifiers used in these experiments are described as reverse demulsifiers by Witco Chemical.

TABLE 3 - 4

EMULSION CHARACTERIZATION RESULTS

PARAMETER	HIGH OIL CONTENT EMULSIONS P33	P35	P1-11	INITIAL CHARAC.	FINAL CHARAC.	TYPICAL EMULSION
pH	8.16	8.25	8.28	8.01	7.75	6.8 - 8.6
TOC mg/L	16568	24265	703	1560	880	NA
FOC mg/L	359	1169	565	260	165	NA
TDS mg/L	333 *	833 *	1595 *	716	784	844 - 31349
TDO mg/L	NA	NA	NA	124	282	NA
TSO mg/L	26900	22300	5500	1580	1602	96 - 2000
TSS mg/L	450	17216	214	40	34	68 - 447
Naphthenic Acid mg/L	NA	NA	NA	60	NA	NA
SILICA total mg/L	<100	<120	<100	100	56.14	8 - 380
diss. mg/L	91	77	84	100	82.16	NA
Calculated Hardness as $CaCO_3$ mg/L	12.36	11.37	53.27	3.93	3.78	41 - 890
Turbidity (NTU)	34000	44000	190	1150	975	NA
Anions						
Fluoride mg/L	5.2	6.2	4.2	4.13	3.4	NA
Chloride mg/L	360	240	410	142	148	199 - 13348
Nitrite mg/L	<1	<1	<1	NA	<4	NA
Phosphate mg/L	<.01	<.01	<.01	<10	<1	NA
Nitrate mg/L	<.4	<.4	<.04	<.5	<1	0 - 0.3
Sulphate mg/L	119	77	147	66.7	58	22 - 525
Cations						
Beryllium mg/L	0.015	0.006	<.005	<.0005	<.01	NA
Molybdenum mg/L	1.1	1.33	0.178	<.2	0.08	NA
Calcium mg/L	3.4	3.3	17	1.16	1.1	48 - 290
Vanadium mg/L	20.9	19.5	1.15	0.012	0.32	NA
Aluminum mg/L	1.58	1.00	1.06	1.22	0.96	NA
Magnesium mg/L	0.94	0.76	2.64	0.25	0.25	5 - 156
Barium mg/L	0.07	0.063	0.23	0.011	0.13	NA
Potasium mg/L	7.6	6.3	19.9	6	6.5	NA
Strontium mg/L	0.115	0.093	0.78	0.010	<.01	NA
Sodium mg/L	300	265	385	245	225	215 - 11007
Zinc mg/L	0.13	0.12	0.28	<.05	0.05	NA
Cadmium mg/L	0.02	0.02	0.04	<.01	<.01	NA
Manganese mg/L	0.12	0.03	0.83	0.02	0.05	NA
Cobalt mg/L	0.06	0.02	0.02	<.05	<.05	NA
Copper mg/L	0.04	0.01	0.31	0.073	0.05	NA
Silver mg/L	0.02	<.03	<.03	<.005	<.02	NA
Iron mg/L	1.44	0.93	2.4	7.92	71	0.1 - 1.0
Lead mg/L	0.12	0.04	0.18	<.05	<.02	NA
Chromium mg/L	0.3	0.23	0.14	<.01	0.06	NA
Nickel mg/L	6.5	6.3	0.62	<.05	0.13	NA

TOC	total organic carbon	TSO	total suspended oil
FOC	filtered organic carbo	TSS	total suspended solids
TDS	total dissolved solids	TDO	total dissolved oil
*	the values include TDS+TDO		Hardness is calculated from Ca & Mg

NA - DATA NOT AVAILABLE

TYPICAL EMULSION CHARACTERIZATION DATA IS FROM CH2M HILL (1982)

The results for the standard emulsion show an extremely low hardness of 4 mg/L. Typical hardness values for emulsions are reported in the range between 41 and 890 mg/L (CH_2M HILL, 1982). The hardness of the three high oil content samples were 11, 12, and 53 mg/L. Some variability at the Texaco site is apparent but all of the values are below or in the lower end of the range of typical values.

Other components in the standard emulsion sample fall within or near the typical range. The three high oil samples show considerable variability in TOC, TDS, TSO, TSS, and turbidity.

The iron contents of the three higher oil samples were 0.93, 1.44, and 2.4 mg/L. The iron contents determined in the initial and final analyses of the standard emulsion were 8 mg/L and 71 mg/L respectively. The nine fold increase in iron between the time of the initial and final analyses is believed to be the result of iron leaching out from the metal storage drums. It is likely that the 8 mg/L determined initially is also somewhat elevated as a result of leaching from the metal drum. The high oil content samples were shipped in plastic containers and analyzed soon after they arrived. The iron content determined should accurately represent the iron present in the emulsion when it was sampled. The stability of the standard emulsion appeared unaffected over time with the increasing concentration of iron.

Oil and grease and ethyl acetate extractable surfactant yielded a total of 1705 mg/L of oil and grease and 218 mg/L of surfactant. The oil and grease value, which was determined by freon extraction is very close to the 1620 mg/L obtained by the method used for TSO and TDO in previous analyses.

3.1.4 GC and IR Analysis

Gas chromatographic (GC-FID) analysis of the pH 2 emulsion sample extracted with CH_2Cl_2 showed no distinctive peaks and therefore conclusions about the content of the sample are difficult to make.

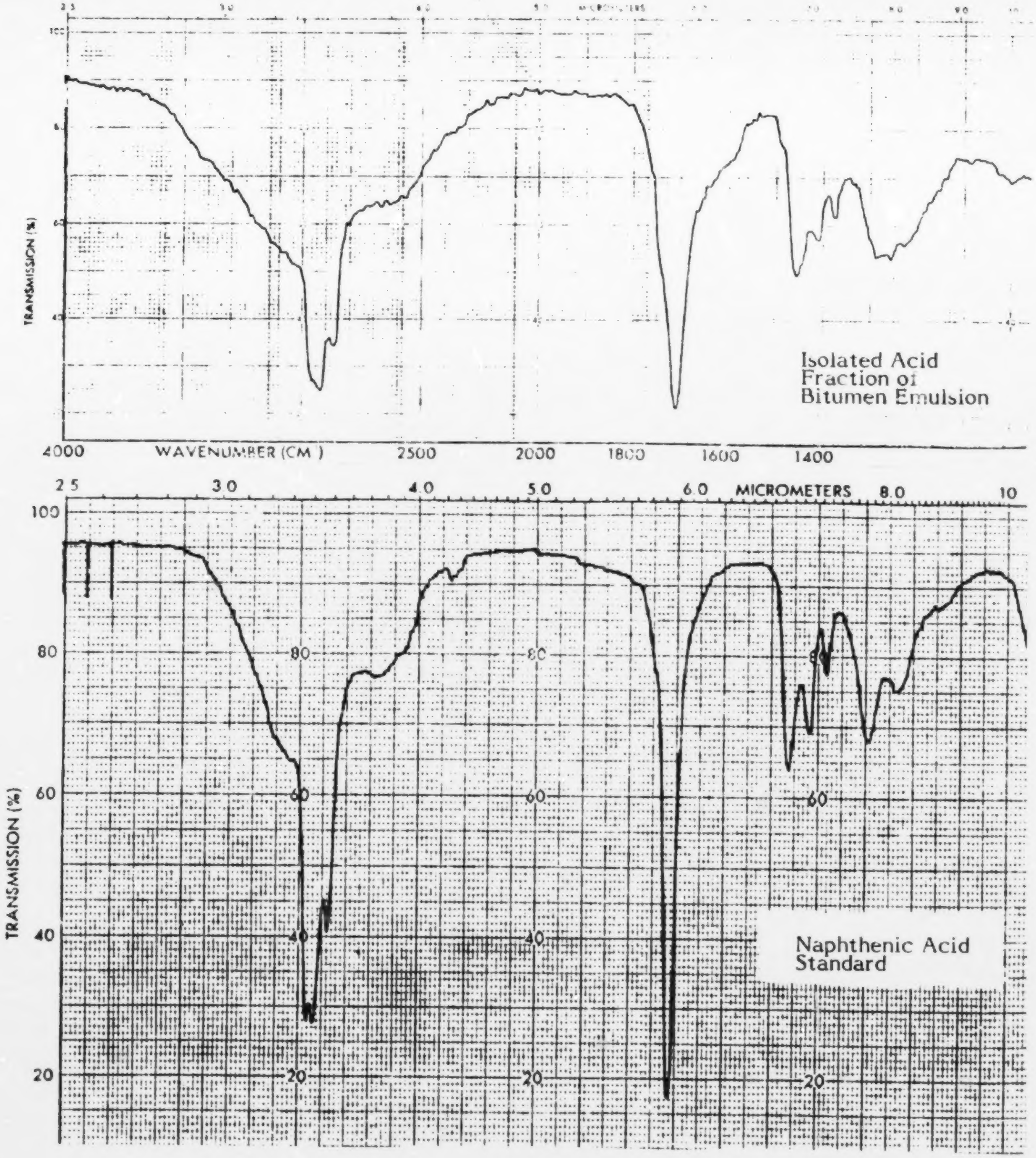
Infra-red spectrophotometric analysis (IR) of the residue on the IR plate after evaporation of solvent gave a spectrum with very broad adsorptions. A definite O-H stretch at 3800 - 2800 cm^{-1} is present. Absorptions at 2000 cm^{-1} and 1760 - 1500 cm^{-1} are also present but unambiguous assignment is not possible. A wide band between 1760 - 1500 cm^{-1} is felt to be carbonyl. A saturated aliphatic C-H stretch is observed at 2970 cm^{-1}. No other adsorption bands are visible. The IR spectra for these tests are presented in Appendix B.

GC analysis of the aqueous supernatant of the broken bitumen emulsion at pH 2 showed a large hump commonly associated with naphthenic acids. The size of this hump indicated a concentration of 13 mg/L of naphthenic acid in the sample. IR analysis showed a strong OH stretch associated with organic acids at 1700 - 1740 cm^{-1} and peaks at 1450, 1410, and 1370 cm^{-1}. All these peaks are consistent with the presence of naphthenic acid in the sample.

The third emulsion sample was back extracted to eliminate interference from other non acidic organic components. IR analysis of the back-extracted sample of bitumen emulsion showed an exact match to a naphthenic acid standard. Figure 3-1 shows the IR scan for the isolated acid fraction of the emulsion and for a naphthenic acid standard. GC analysis showed the presence of 60 mg/L of naphthenic acid.

FIGURE 3 - 1

INFRARED SCAN OF ISOLATED ACID FRACTION OF BITUMEN EMULSION AND NAPHTHENIC ACID STANDARD



It is beleived that the naphthenic acids present in the emulsion are primarily responsible for maintaining the very high stability of the emulsion. When the pH of the emulsion is lowered, aqueous phase solubility is lowered and much of the naphthenic acid passes into the oil phase and is no longer active as a surfactant. This is illustrated by the fact that the aqueous supernatant of the pH 2 emulsion contained only 13 mg/L of naphthenic acid, considerably less than the 60 mg/L present in the isolated acid fraction of the bitumen emulsion.

3.2 **Membrane Prescreening Test**

3.2.1 Pure Water Flux Tests

A summary of the results of the pure water flux tests on the 20 membranes is shown in Table 3-5. Complete results presented in Appendix C. The results give an indication of the flux rates obtainable with each membrane. The polysulfone (PS) and CHP membranes consistently gave the highest flux rates. Flux rates for TFC membranes increased as the pore sizes of the membranes increased. The cellulose acetate (CA) membrane 50K has a much lower flux than the CA - 0K or 20K membrane. The lower flux of the larger pore membrane is believed to be the result of a different backing material used with CA - 50K restricting the flux rate.

3.2.2 Salt Water Flux Test

A summary of the results of the salt water flux test is presented in Table 3-6. Complete results are presented in Appendix C. These results give a furtner indication of potential flux rates

TABLE 3 - 5

RESULTS OF PURE WATER FLUX TESTS

MEMBRANE	AVERAGE FLUX RATE M3/M2 D	MEMBRANE	AVERAGE FLUX RATE M3/M2 D
TFC G5	0.31	PS-0K	12.29
G10	1.24	20K	16.62
G20	3.10	50K	15.13
G50	11.26	E500	16.46
CA-0K	3.36	CELLULOSE 12	0.47
20K	6.04	13	3.23
50K	1.72	23	1.72
UF	1.14	RC-51	0.20
C.H.P. Cat	19.50	PVDF	5.91
B2	11.31		
B3	13.12		

Operating Temperature = 25 deg. C
Operating Pressure = 100 psi (712 KPa))

TABLE 3 - 6

RESULTS OF SALT WATER FLUX TESTS

MEMBRANE	AVERAGE FLUX RATE M3/M2 D	% REJ	MEMBRANE	AVERAGE FLUX RATE M3/M2 D	% REJ
TFC G5	0.28	71.43	PS-OK	14.28	7.14
G10	0.96	72.81	20K	14.44	9.70
G20	2.67	50.38	50K	49.32	11.82
G50	1.03	72.67	E500	13.96	13.96
CA-OK	6.41	1.70	CELLULOSE 12	0.40	6.70
20K	5.66	0.00	13	0.19	18.81
50K	1.60	48.57	23	1.22	5.02
UF	11.28	9.72	RC-51	0.17	20.58
C.H.P. Cat	15.43	74.35	PVDF	4.61	10.02
B2	11.56	88.91			
B3	12.40	26.59			

Operating Temperature = 25 deg. C
Operating Pressure = 100 psi (712 KPa)

through each membrane and an indication of the percent rejection of monovalent ions each membrane can achieve. As in the pure water flux tests the flux rates for PS and CHP membranes were consistently higher than for all other membranes. The salt rejections for the PS membranes were however consistently much poorer than for the CHP membranes. Three membranes were dropped at this point because of consistently poor performance during these tests. The membranes not included in further work were TFC-G5, CA-UF, and PS-OK.

3.2.3 Emulsion Prescreening Test

Figures 3-2, 3-3, and 3-4 show the flux rate results of the emulsion prescreening tests. Table 3-7 shows the efficiency of each membrane in rejecting the TOC present in the bitumen emulsion during the prescreening test. Complete results of these tests are presented in Appendix C.

The pure water and salt water tests yield information on membrane characteristics but the bitumen prescreening tests results were the key in identifying optimum membranes for this application. The final selection of membranes was made by the scientific authority after an evaluation of the results had been made.

The three TFC membranes tested all exhibited very good TOC rejections of between 98 - 99%. The flux rate for the G10 membrane was initially lower than the other TFC membranes but showed virtually no decrease with time. Membrane G50 had a higher flux rate than the G10 but was decreasing at a slow but steady rate. Membrane G20 was initially higher than both the G10 and G50 membranes but by the end of 3 1/2 hours had decreased to below that for the other two membranes.

FIGURE 3 - 2

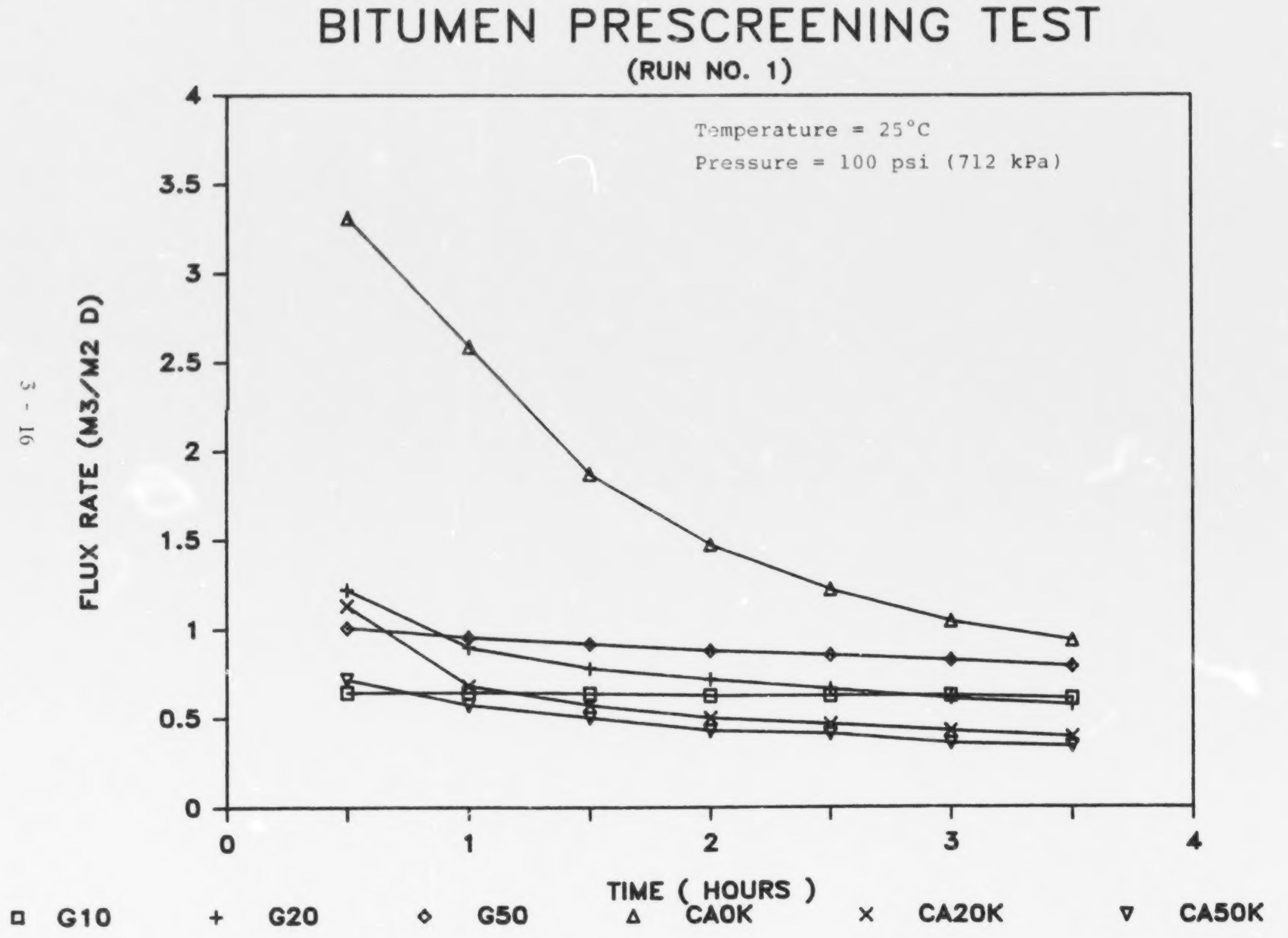

BITUMEN PRESCREENING TEST
(RUN NO. 1)
Temperature = 25°C
Pressure = 100 psi (712 kPa)
FLUX RATE (M3/M2 D)
0
0.5
1
1.5
2
2.5
3
3.5
4
0
1
2
3
4
TIME (HOURS)
□ G10
+ G20
◇ G50
△ CA0K
× CA20K
▽ CA50K

FIGURE 3 - 3

BITUMEN PRESCREENING TEST

(RUN NO. 2)

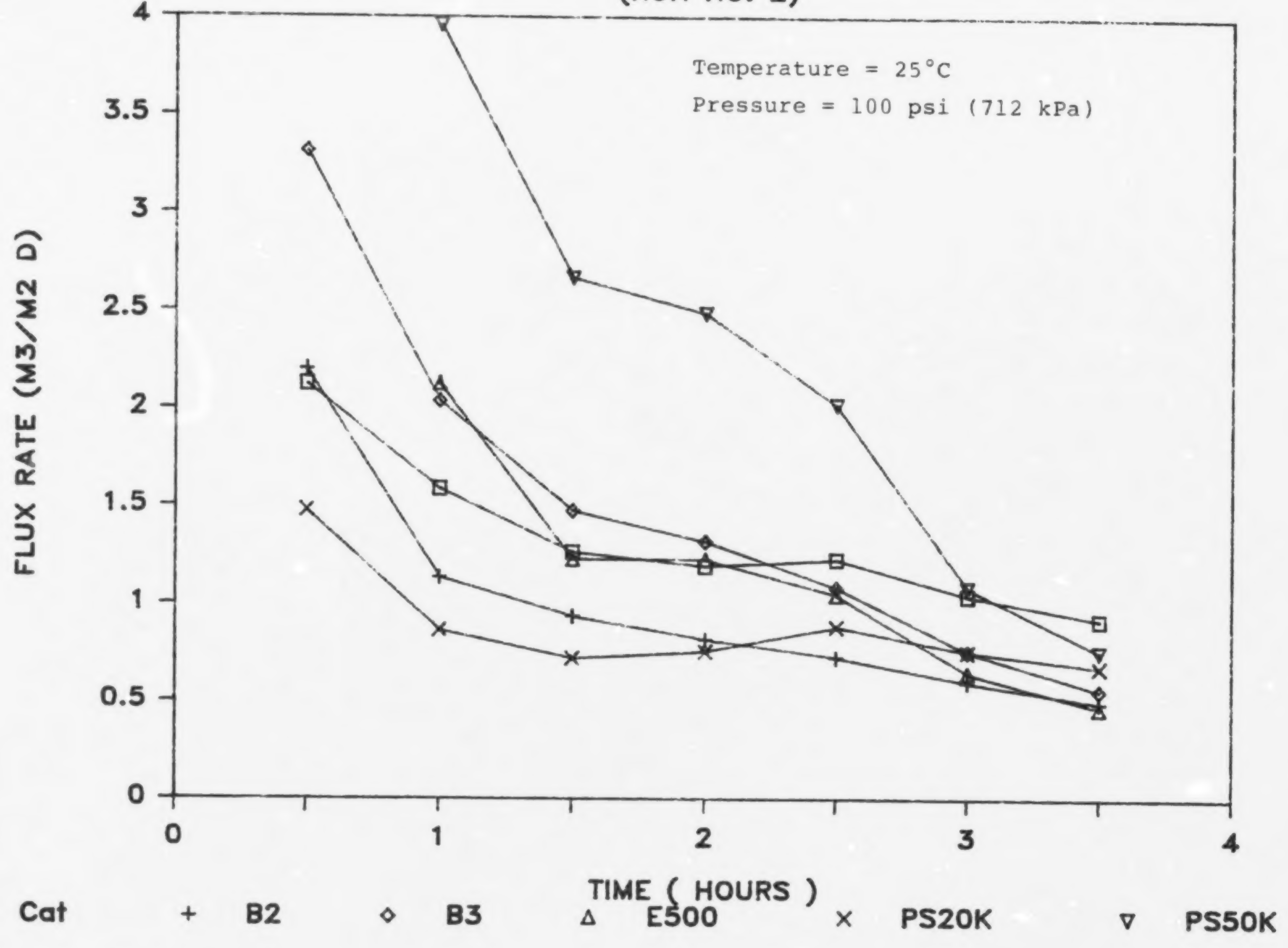


FIGURE 3 - 4

BITUMEN PRESCREENING TEST

(RUN NO. 3)

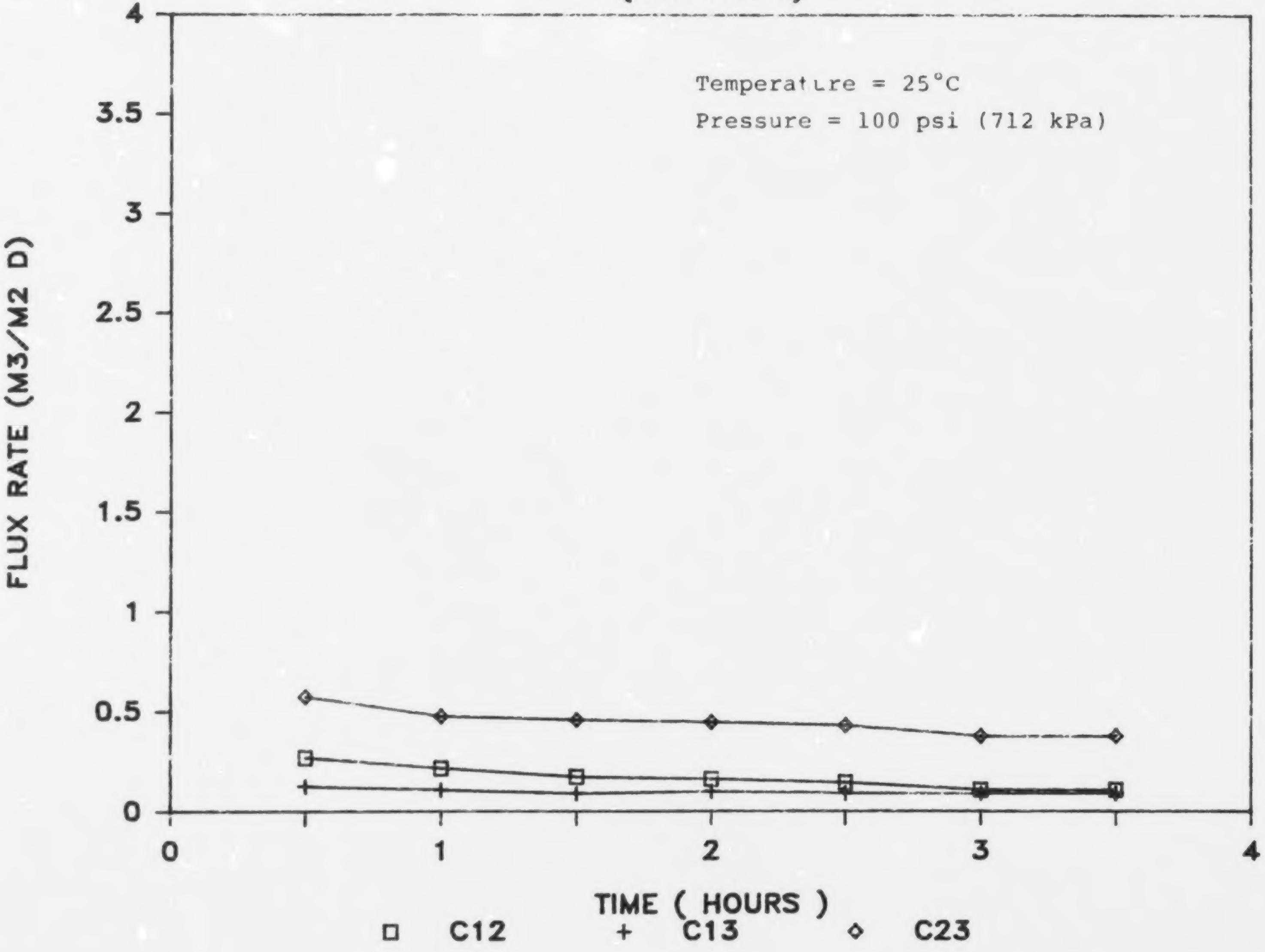


TABLE 3 - 7

MEMBRANE EFFICIENCY DURING BITUMEN PRESCREENING TESTS

MEMBRANE	(AVERAGE) % REJECTION TOC	MEMBRANE	(AVERAGE) % REJECTION TOC
TFC G10	99.34	PS E500	96.82
TFC G20	97.94	PS 20K	94.34
TFC G50	98.17	PS 50K	96.87
CA OK	96.17	CELLULOSE 12	91.78
CA 20K	93.66	CELLULOSE 13	88.33
CA 50K	93.66	CELLULOSE 23	94.77
CHP Cat	97.48		
CHP B2	48.43		
CHP B3	97.61		

Operating Temperature = 25 deg. C
Operating Pressure = 100 psi (712 KPa)

Based on the steady flux for G10 it was chosen as the best TFC membrane for further work.

The three cellulose acetate membranes, CA-0K, CA20K, and CA50K exhibited TOC rejections of 96%, 94%, and 94% respectively. The flux rate for CA-0K was initially high but decreased very rapidly during the 3 1/2 hour run. The flux rates for CA20K and CA50K were both low but relatively stable during the run. The CA20K membrane was chosen as a result of its stable flux.

CHP membranes Cat and B3 had rejections of TOC of 97.5% and 97.6%. The flux rate for B3 was initially very high but decreased rapidly during the test. The CHP-Cat membrane showed some fluctuation in flux rate but was relatively stable with time. The flux for Cat was consistently higher than that of B2. The rejection for CHP B2 was only 48.4%. The flux rate for CHP B2 declined steadily during the test. Cat was chosen as a result of its high stable flux and good rejection of TOC.

Polysulfone (PS) membranes E500, 20K, and 50K obtained TOC rejections of 96.8%, 94.3%, and 96.9% respectively. The flux rate for PS50K was initially very high but decreased rapidly over the 3 1/2 hours of the test. The E500 membrane had a high flux which gradually decreased during the test and the PS20K had a lower more stable flux. The E500 membrane was chosen over the PS20K membrane even though the flux rate of the PS20K was stable because of the more ready availability of the E500 membrane in a variety of configurations.

The three cellulose membranes C12, C13, and C23 had rejections of 91.8%, 88.3%, and 94.8% respectively. The flux rates for all three membranes were stable but low relative to the other membranes. As a result of the low flux rates of all three cellulose membranes, they were dropped from the experimental program at this time. The cellulose membrane RC-51, which was not available at the time of the prescreening tests, was used to replace the other cellulose membranes.

The polyvinyl difluoride (PVDF) membrane was not tested because only one such membrane was available for testing.

3.2.4 Polyethelene Glycol (PEG) Test

The results of the PEG test were used to characterize the selected membranes with respect to their molecular weight cut offs. A summary of the results of the PEG test showing the percent rejection of TOC and the estimated molecular weight cut offs for the membranes is shown in Table 3-8. Appendix C contains the complete data for the test.

The molecular weight cut offs were estimated by comparing the percent rejection of TOC for each membrane with percent rejections of TOC for membranes of known molecular weight cut offs. Four of the membranes, CA-20K, PS-E500, CHP-CAT, and PVDF had percent rejections between 5% and 15% suggesting molecular weight cut offs between 15-20K. The TFC-G10 membrane and the Cellulose RC51 had much higher rejections of 82% and 88% suggesting molecular weight cut offs between 6-10K.

TABLE 3 - 8

ESTIMATED MOLECULAR WEIGHT CUT OFF BASED ON THE PEG TEST

MEMBRANE	PEG % REJECTION OF TOC	ESTIMATED MOLECULAR WEIGHT CUT OFF
CA-S-20K	5%	15 - 20K
PS-E500	8%	15 - 20K
TFC-G10	82%	6 - 10K
CHP-Cat	13%	15 - 20K
C-RC51	88%	6 - 10K
PVDF	8%	15 - 20K

Operating Temperature = 25 deg. C
Operating Pressure = 100 psi (712 KPa)

3.3 **Membrane Screening Tests**

The flux rates for the eight screening test runs are shown in Figures 3-5 to 3-12. Complete results of the screening tests are presented in Appendix D.

The flux rates for the cellulose membrane during the first six runs were consistently very poor. During the seventh run the cellulose membrane was replaced with an additional sample of CHP membrane. The flux rates for the other membranes were similar during the runs but the cellulose acetate membrane tended to have slightly higher flux rates than the others. Of the remaining four membranes no one membrane in particular consistently out performed the others.

In order to evaluate whether temperature and pH had a significant effect on the flux rate for each membrane two statistical tests were applied to the results. The flux rate at a 20% volume reduction for each membrane during each run were tabulated. This data is presented in Table 3-9. A correlation test and a multiple linear regression were applied to the data for each membrane using temperature and pH as the independent variables and flux rate at 20% volume reduction as the dependent variable.

Table 3-10 shows the T-test results from the linear regression test for the pH and temperature coefficients. Complete results of the correlation test and multiple linear regression are presented in Appendix D. The higher the T-test result on a particular coefficient the greater the likelihood that the factor does have a significant effect. In order to have a 95% probability that a factor is significant with five degrees of freedom the T-test value must be

FIGURE 3 - 5

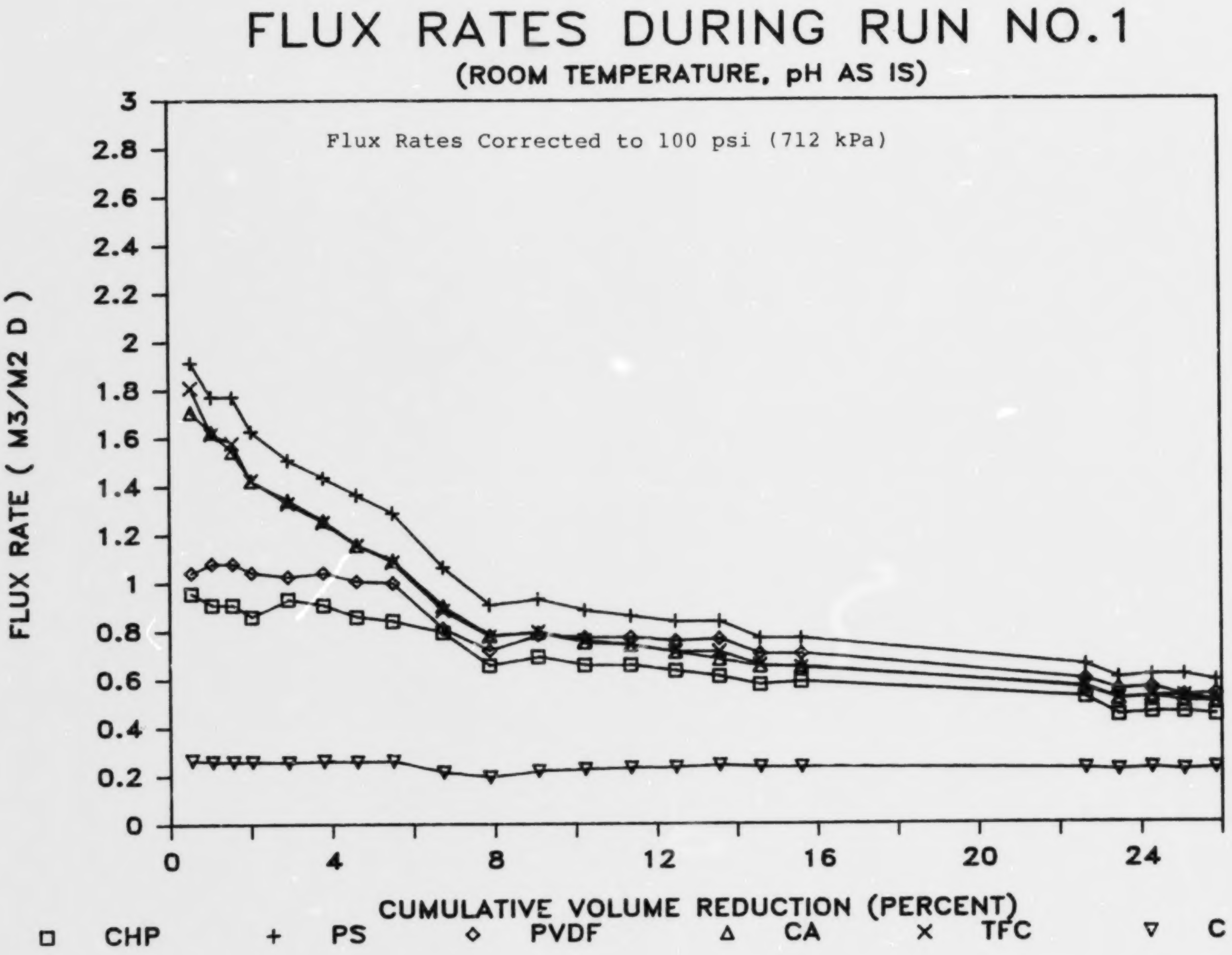

FLUX RATES DURING RUN NO.1
(ROOM TEMPERATURE, pH AS IS)
Flux Rates Corrected to 100 psi (712 kPa)
FLUX RATE (M3/M2 D)
3
2.8
2.6
2.4
2.2
2
1.8
1.6
1.4
1.2
1
0.8
0.6
0.4
0.2
0
0
4
8
12
16
20
24
CUMULATIVE VOLUME REDUCTION (PERCENT)
□ CHP
+ PS
◇ PVDF
△ CA
× TFC
▽ C

FIGURE 3 - 6

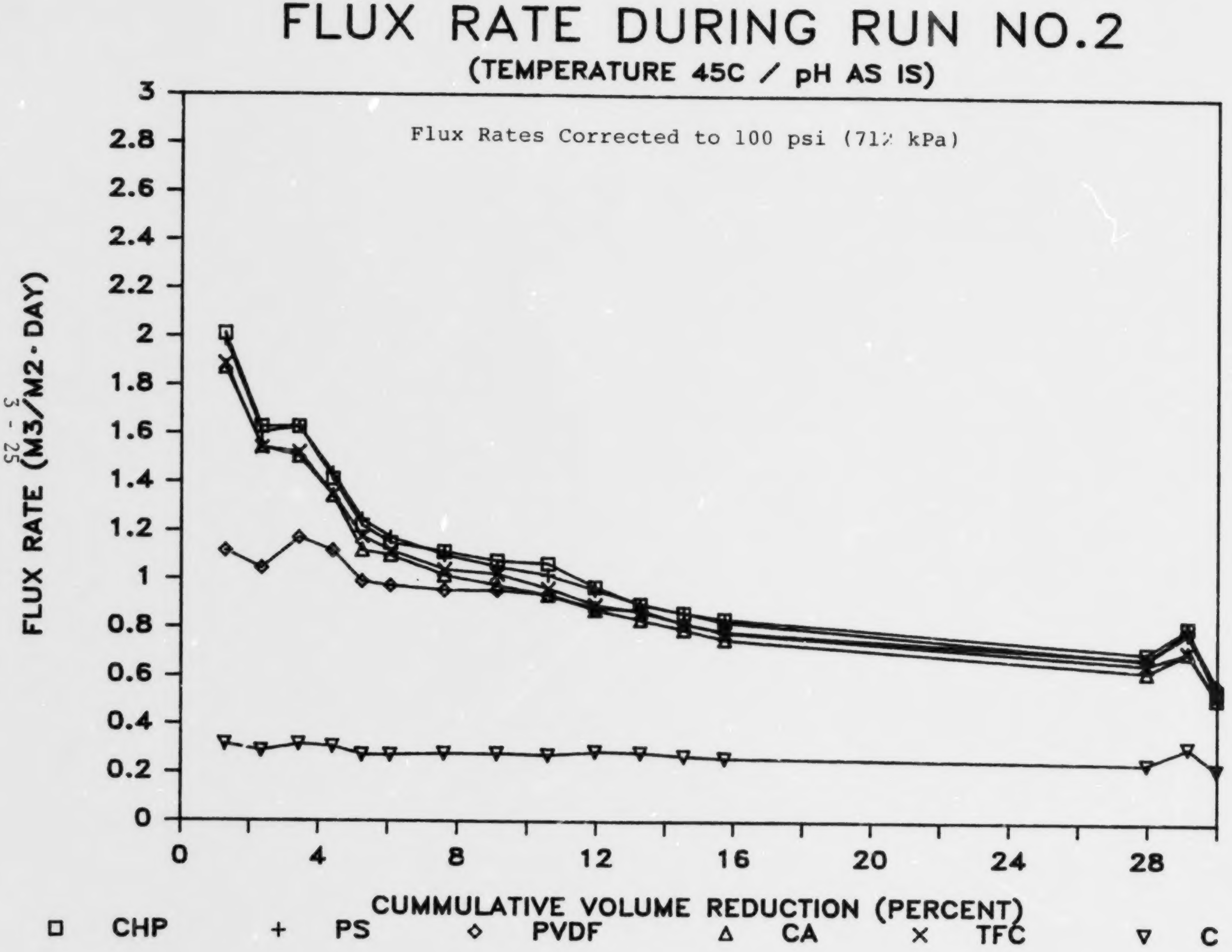

FLUX RATE DURING RUN NO.2
(TEMPERATURE 45C / pH AS IS)
Flux Rates Corrected to 100 psi (71% kPa)
FLUX RATE (M3/M2·DAY)
0
0.2
0.4
0.6
0.8
1
1.2
1.4
1.6
1.8
2
2.2
2.4
2.6
2.8
3
CUMMULATIVE VOLUME REDUCTION (PERCENT)
0
4
8
12
16
20
24
28
□ CHP
+ PS
◇ PVDF
△ CA
× TFC
▽ C

FIGURE 3 - 7

FLUX RATE DURING RUN NO.3

(ROOM TEMPERATURE, pH 5.5)

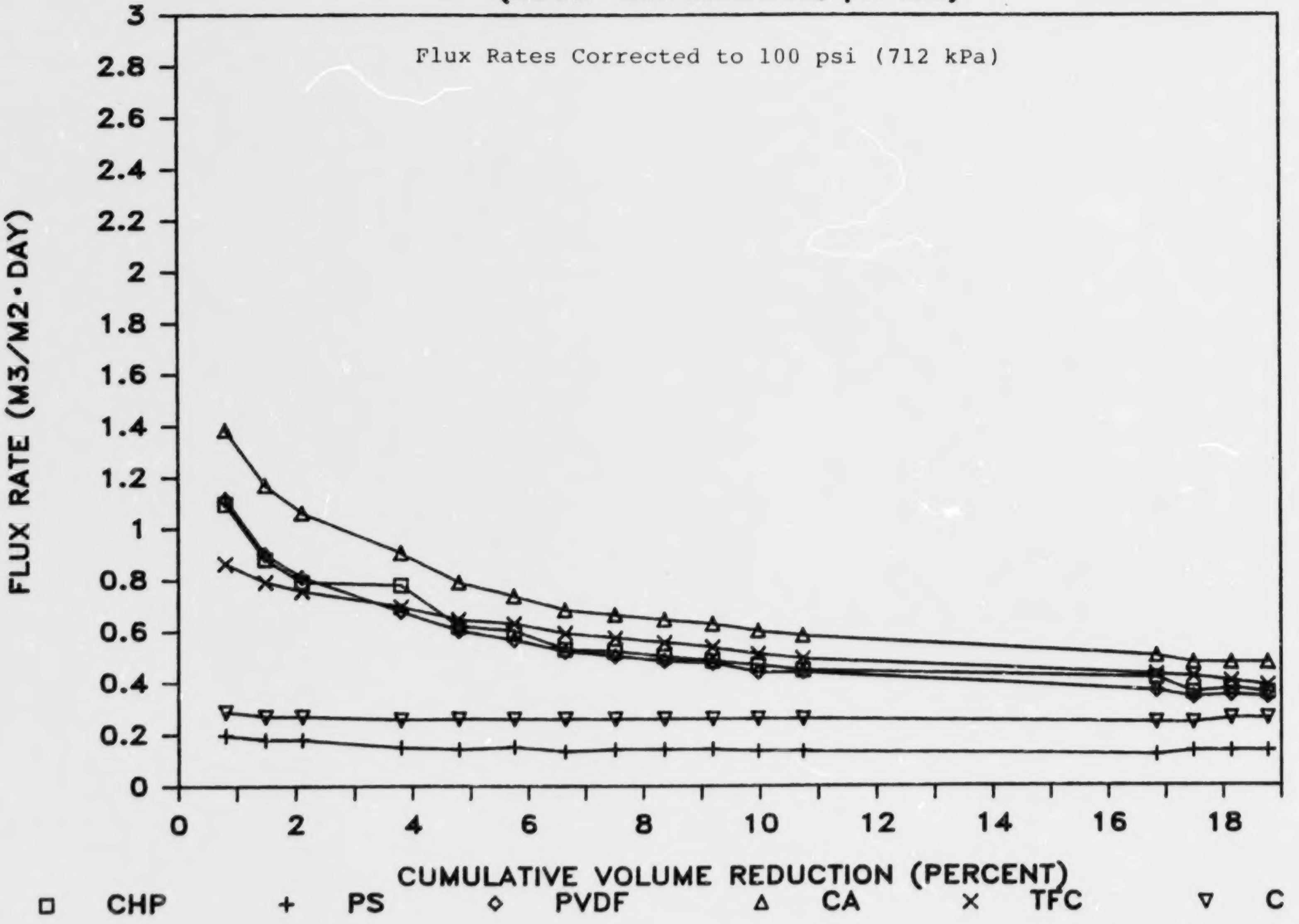


FIGURE 3 - 8

FLUX RATE DURING RUN NO.4

(TEMPERATURE 45C / pH 5.5)

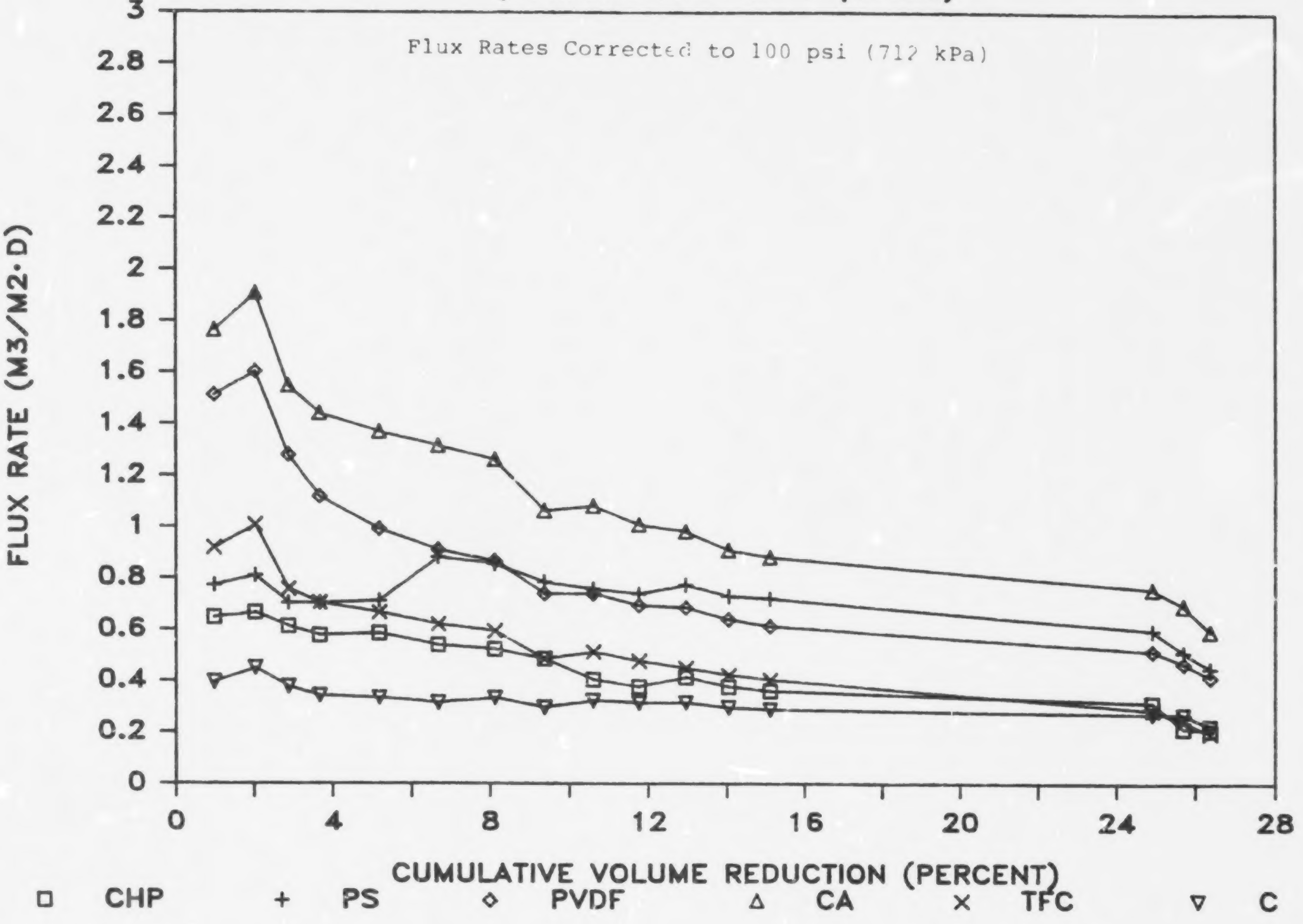


FIGURE 3 - 9

FLUX RATE DURING RUN NO.5

(ROOM TEMPERATURE / pH 9.5)

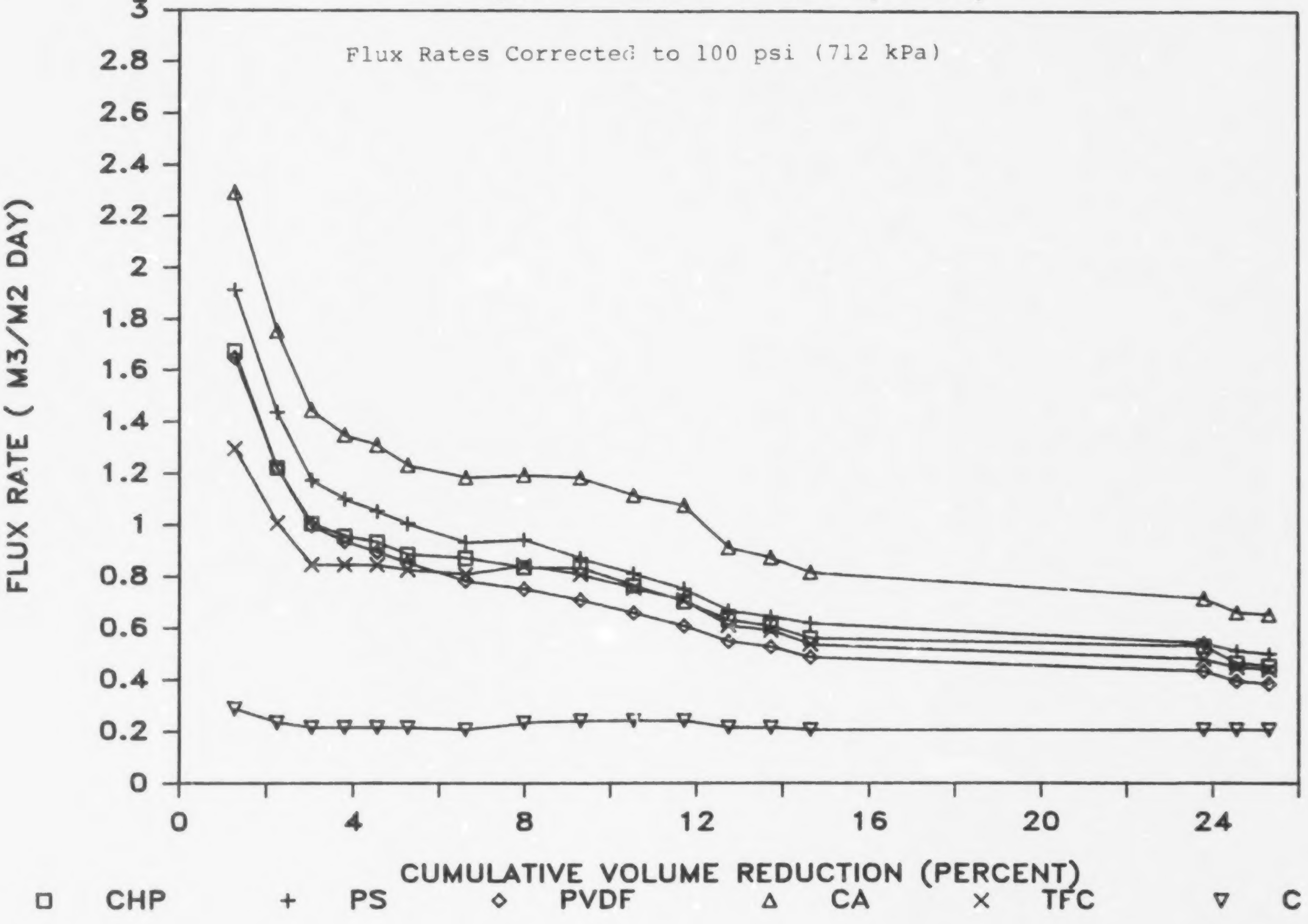


FIGURE 3 - 10

FLUX RATE DURING RUN NO.6

(TEMPERATURE 45 C / pH 9.5)

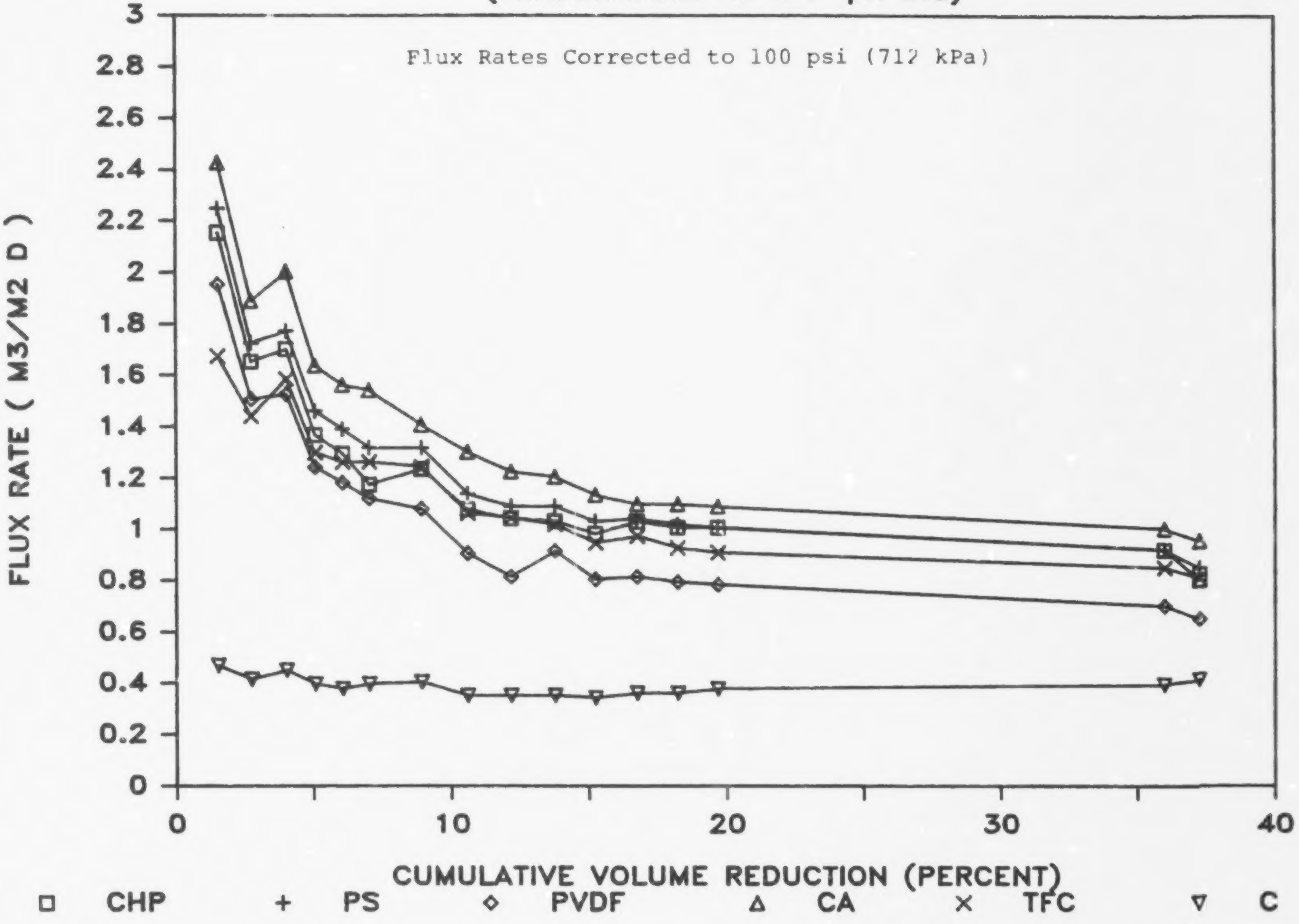


FIGURE 3 - 11

FLUX RATES DURING RUN NO.7

(TEMPERATURE 65 C / pH AS IS)

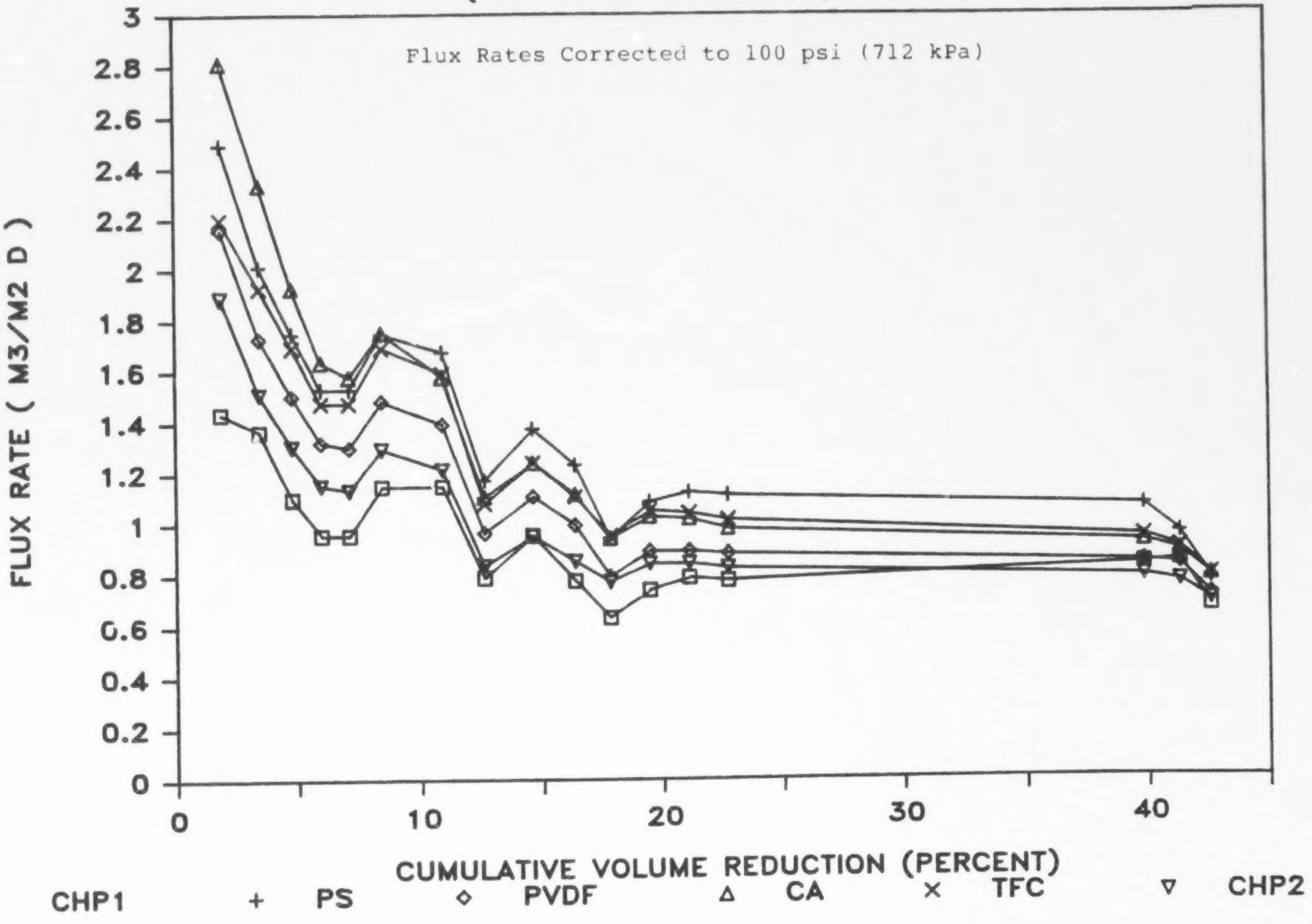


FIGURE 3 - 12

FLUX RATES DURING RUN NO.8

(TEMPERATURE 65 C / pH AS IS)

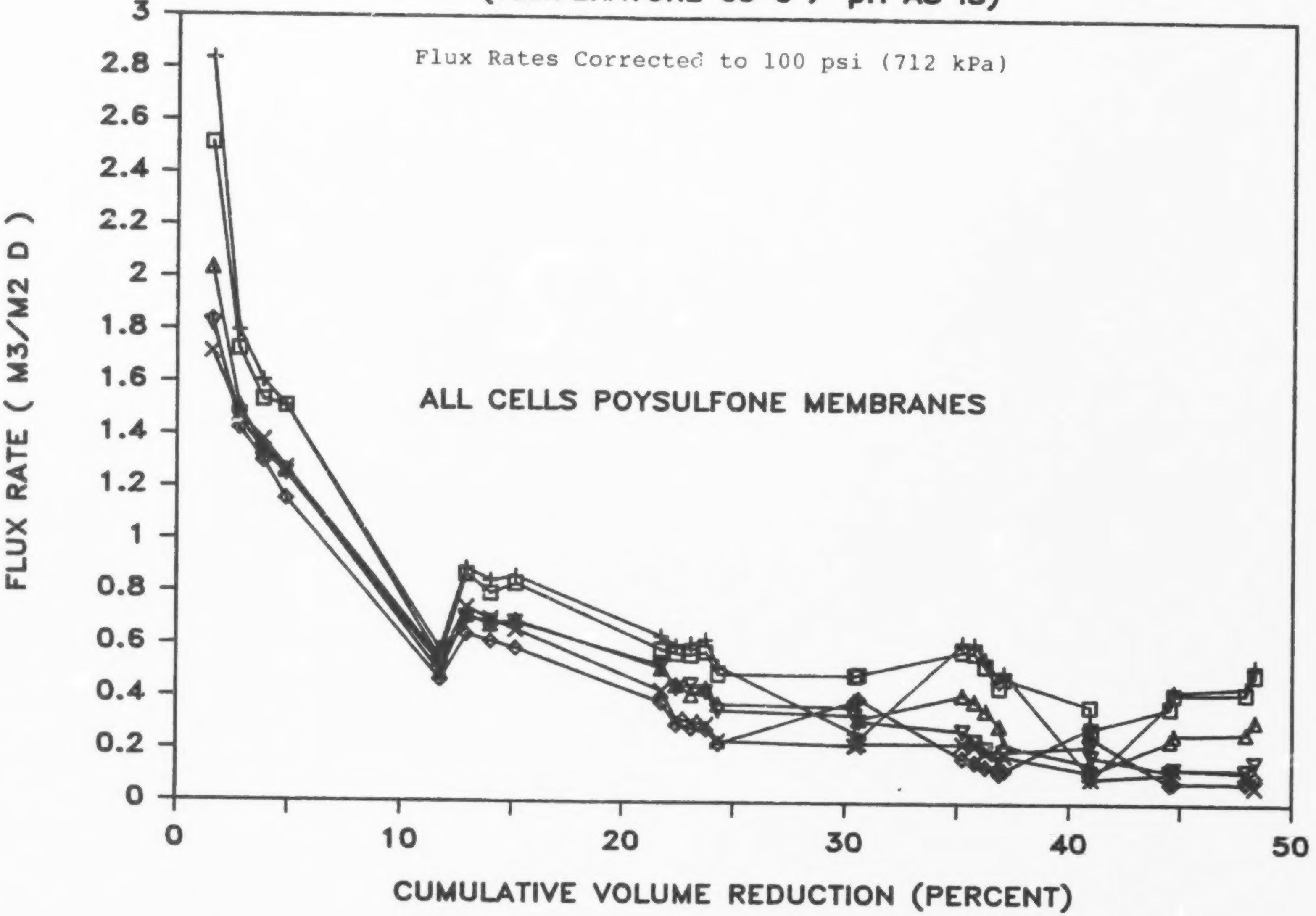


TABLE 3 - 9

FLUX RATES AT 20 % VOLUME REDUCTION

RUN NO.	CHP FLUX RATE M3/M2 D	PS FLUX RATE M3/M2 D	TFC FLUX RATE M3/M2 D	C FLUX RATE M3/M2 D	CA FLUX RATE M3/M2 D	PVDF FLUX RATE M3/M2 D
1	0.55	0.70	0.63	0.22	0.59	0.59
2	0.79	0.78	0.75	0.25	0.74	0.71
3	0.48	0.18	0.39	0.26	0.51	0.39
4	0.45	0.88	0.35	0.28	0.88	0.65
5	0.55	0.57	0.50	0.21	0.76	0.45
6	0.99	0.99	0.90	0.38	1.07	0.77
7	0.81	1.10	1.05	ND	1.03	0.89
MEAN	0.66	0.74	0.65	-	0.80	0.64

RUN 1 pH AS IS, ROOM TEMPERATURE (25C)
2 pH AS IS, TEMPERATURE (45C)
3 pH 5.5, ROOM TEMPERATURE (25C)
4 pH 5.5, TEMPERATURE (45C)
5 pH 9.5, ROOM TEMPERATURE (25C)
6 pH 9.5, TEMPERATURE (45C)
7 pH AS IS, TEMPERATURE (65C)

ND - not done

All Flux Rates are Pressure Corrected to 100 psi (712 KPa)

TABLE 3 - 10

SIGNIFICANCE TEST OF pH AND TEMPERATURE COEFFICIENTS IN MULTIPLE LINEAR REGRESSION

MEMBRANE	T-TEST ON pH COEFFICIENT	T-TEST ON TEMP CEFFICIENT
CHP	2.34	2.28
PS	1.45	3.56
TFC	2.08	2.85
CA	2.03	3.82
PVDF	1.24	5.42
CELLULOSE	0.43	1.55

greater than 2.02. On this basis it is apparent that temperature does have a significant effect on the flux rates.

The cellulose T-test result for temperature is less significant partially because no run was conducted at 65°C with this membrane. Figure 3-13 shows the flux rates for each membrane during runs at different temperatures.

The results showed consistently positive correlation between flux rate and increasing pH. The T-test data also shows pH to be a significant factor in determining flux rate for some of the membranes. It is possible that by increasing the pH of the emulsion, the emulsion stability is enhanced, and membrane fouling tends to decrease. Although the increase in flux with increasing pH is not as significant as increases in flux with temperature its effect should be noted. Figure 3-14 shows the flux rates for each membrane at different pH values.

Table 3-11 shows a summary of the permeate qualities generated during the screening tests. This data shows generally poor rejection of components by the cellulose membrane, similar rejections by the CA, PS, CHP, and PVDF membranes, and generally better rejections by the TFC membrane.

Table 3-12 shows the range of percent rejections for different components obtained for each membrane during the first seven runs. The rejection of conductivity by the cellulose membrane was poor ranging between 0-14%. The CA, PS, CHP and PVDF membranes were better with rejections ranging between 25-53%, 31-65%, 33-68%, and 36-58% respectively. The TFC membrane was still better with percent rejections between 63-81%. No trends in rejection of conductivity

FIGURE 3 - 13

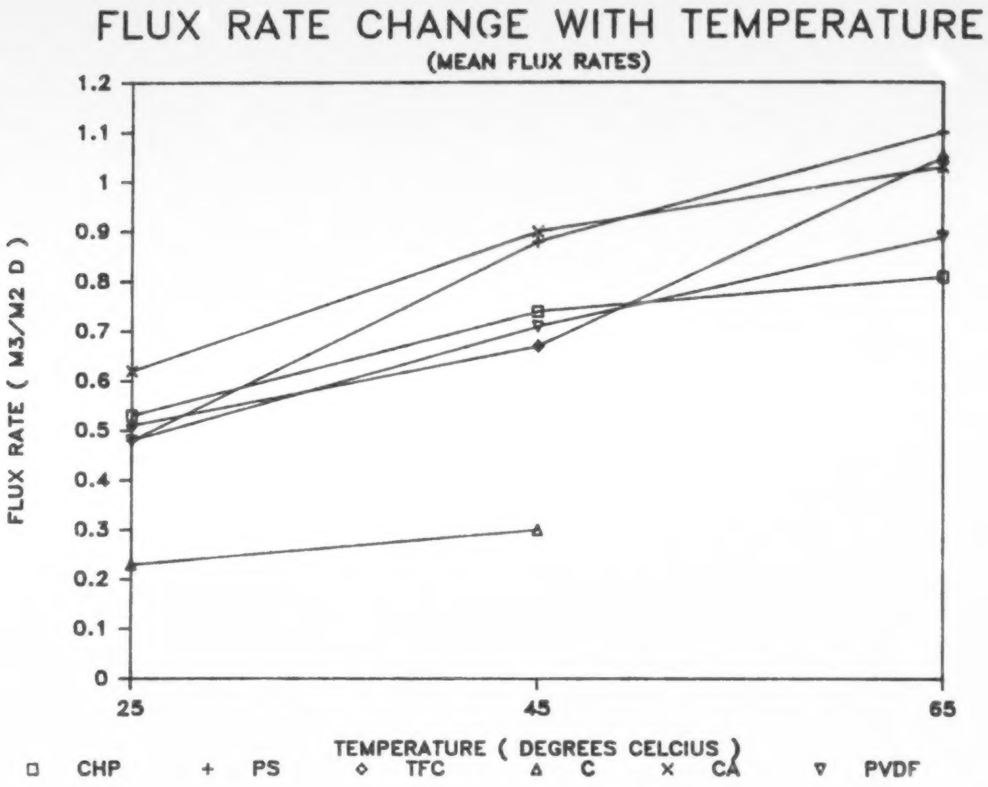


FIGURE 3 - 14

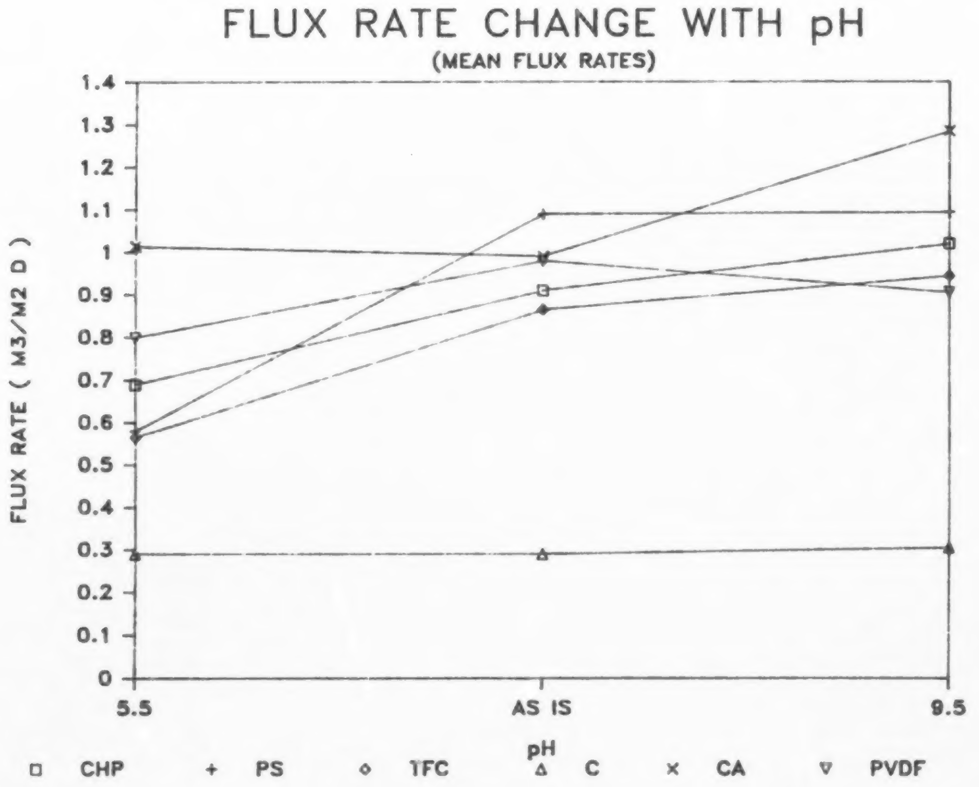


TABLE 3 - 11

PERMEATE QUALITY DURING SCREENING TESTS

	CONDUCTIVITY (uHmos)	TURBIDITY (NTU)	TOC (mg/L)	TDS (mg/L)	HARDNESS (mg/L)	SILICA (mg/L)
FEED	913 - 1210	1090 - 2410	1025 - 2919	980 - 1155	3.1 - 36.1	61 - 85
PERMEATES						
CELLULOSE	850 - 1075	0.8 - 1.2	42 - 140	605 - 845	0.6 - 28	40 - 69
CA	470 - 800	0.4 - 0.9	32 - 88	308 - 561	<0.15 - 7.2	41 - 70
PS	385 - 630	0.4 - 2.2	32 - 66	399 - 475	<0.15 - 8.7	36 - 70
CHP	358 - 640	0.4 - 0.9	21 - 76	362 - 470	<0.15 - 4.5	39 - 68
PVDF	420 - 590	0.3 - 0.7	22 - 67	325 - 430	<0.15 - 4.2	39 - 71
TFC	195 - 375	0.1 - 0.4	8 - 45	160 - 262	<0.15 - 0.75	40 - 64

TABLE 3 - 12

PERCENT REJECTIONS OF DIFFERENT COMPONENTS
DURING MEMBRANE TESTS

	PERCENT REJECTION				
MEMBRANE	CONDUCTIVITY	TOC	TDS	HARDNESS	SiO2
CELLULOSE	0 - 14	91 - 96	19 - 44	23 - 86	19 - 43
CA	25 - 53	94 - 98	43 - 73	67 - 99	18 - 39
PS	31 - 65	95 - 98	51 - 64	82 - 99	18 - 42
CHP	33 - 68	94 - 98	52 - 68	81 - 99	20 - 42
PVDF	36 - 58	96 - 98	59 - 70	84 - 99	16 -44
TFC	63 - 81	97 - 98	74 - 86	98 - 99	18 - 42

with respect to pH and temperature were observed. All membranes were able to reject virtually all (99.9%) of the turbidity present in the feed. During the eight runs the turbidities of the permeates for all membranes were less than 1 NTU. All membranes rejected virtually all the emulsified oil present in the feed.

The TOC rejections for all membranes were quite high but rejections for the cellulose were generally lower and the rejections for the TFC somewhat higher than for the other membranes. No definite trends for rejection of TOC with respect to temperature were observed but the two lower pH runs tended to have slightly poorer rejections. The rejections of TDS were lowest (19-44%) for the cellulose membrane and highest (74-86%) for the TFC membrane. The others all had rejections between 43-73%. No trends for TDS rejection with respect to pH and temperature were observed.

Rejection of hardness followed a similar pattern as for other components. Rejection were lowest (0-86%) for cellulose and highest (98-99%) for TFC. The high rejections of hardness by the TFC membrane would reduce the load on downstream softening processes in produced water treatment systems. The rejections for CA were slightly lower, 67-99%, than for the PS, CHP and PVDF membranes which ranged from 81-99%. The level of hardness present in the feed increased when the pH of the feed was lowered. The rejection of hardness was also poorer at a pH of 5.5.

The rejections of dissolved silica were very similar for all the membranes ranging from the high teens to the low forties. No trends in the rejection of silica were observed with respect to temperature or pH.

The GC and IR analysis conducted on the permeate from run eight showed naphthenic acids to be present at a concentration of about 7 mg/L. The presence of soluble organic material such as naphthenic acid is not a problem for boiler feeds when, as the results demonstrated, essentially all oil is removed.

Based on the results of these screening tests and discussion with the Scientific Authority three membranes; TFC, PS, and CHP were chosen as the best membranes for further work. The TFC had by far the best rejections of contaminants in the emulsion and maintained a stable and relatively high flux (0.65 m^3/m^2 D mean). The PS membrane had rejections similar to CHP and PVDF membranes and maintained the highest flux (0.74 m^3/m^2 D mean) of the three. The PS membrane is also considerably less expensive than the TFC membrane. The CHP membrane was included because its rejections were good and it maintained a flux rate slightly higher than the PVDF.

The CA membrane maintained the best flux rate of all the membranes (0.80 m^3/m^2 D mean) but had consistently lower rejections of various contaminants than the CHP, PS, TFC, and PVDF membranes. The flux rates and rejection for the cellulose membrane was consistently very poor during the tests.

The results also demonstrated that the operating temperature should be as high as possible in order to maximize the flux rate through the membrane. Temperature appeared to have no impact on the ability of membranes to reject emulsion components.

Increases in pH appeared to have a positive impact on the flux rate of the membranes. Low pH also had a detrimental effect on the removal of hardness by the membranes. If the pH of the water being treated is to have its pH adjusted upwards at some point such as is required for industrial boiler feedwater treatment process it may be desirable to make this upward adjustment before the membrane processing step.

3.4 **Batch Concentration Test**

The flux rate data for the batch concentration test are shown in Table 3-13 and in Figure 3-15. Results from analyses of samples taken during the run are presented in Table 3-14. Complete results are included in Appendix E.

The results showed a rapid drop in flux rate for all three membranes as the cumulative volume reduction increased from 0 to 30%. The flux rates for PS and TFC were stable between 30% and 80% volume reduction near 0.50 and 0.32 m^3/m^2 D respectively. The flux rate for CHP decreased slowly from 0.76 to 0.5 m^3/m^2 D during this time. The flux rates for all three membranes decreased rapidly between 80% and 90% volume reduction. During this period of time the temperature of the emulsion feed which had been stable between 60°C and 65°C decreased to 55°C. The increasing concentration of bitumen in the feed may be partially responsible for the drop in flux rate but the decrease in temperature was likely a significant factor as well.

During the test no sign of any change in emulsion stability was observed. At a 90% volume reduction a sample of the concentrate did not separate even after standing for several hours.

The analytical data from the test demonstrate that the TFC membrane is significantly more efficient in removing TOC than the CHP or PS membranes. At a 90% volume reduction the concentrations of TOC

TABLE 3 - 13

FLUX RATES DURING BATCH CONCENTRATION TEST

TIME (MIN)	CHP FLUX RATE (M3/M2 D)	TFC FLUX RATE (M3/M2 D)	PS FLUX RATE (M3/M2 D)	% VOLUME REDUCTION
0.0				
60.0	3.40	1.40	2.25	2.27
220.2	2.36	1.89	1.95	8.14
1020.0	1.01	0.88	0.83	19.87
1080.0	0.76	0.52	0.35	31.09
1140.0	0.68	0.52	0.37	32.02
1200.0	0.83	0.56	0.39	37.49
1320.0	0.71	0.53	0.34	47.52
1380.0	0.61	0.50	0.33	52.20
1440.0	0.61	0.50	0.31	56.82
1500.0	0.63	0.47	0.31	60.27
2460.0	0.56	0.48	0.30	67.12
2520.0	0.54	0.52	0.33	72.16
2580.0	0.52	0.54	0.33	77.26
2640.0	0.49	0.50	0.31	88.16
2760.0	0.27	0.27	0.15	88.80
2820.0	0.11	0.11	0.06	90.10

Operating Temperature = 65 deg. C
Operating Pressure = 100 psi (712 KPa)

FIGURE 3 - 15

FLUX RATE DURING BATCH CONCENTRATION

(TEMPERATURE 65 C / pH AS IS)

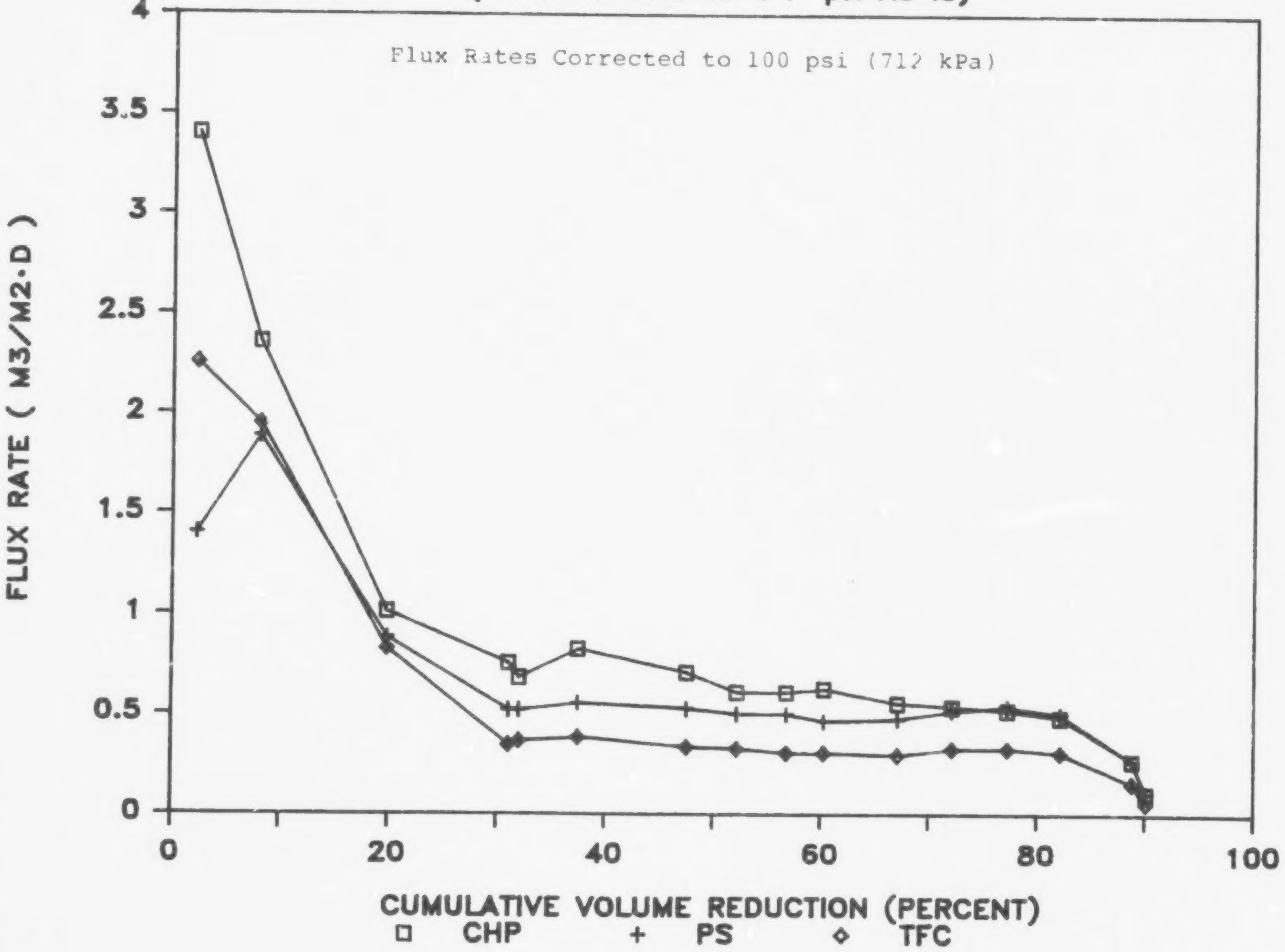


TABLE 3 - 14

BATCH CONCENTRATION RUN ANALYTICAL DATA

PARAMETER	UNIT	FEED	PERMEATE AT 30 % VOLUME REDUCTION			PERMEATE AT 60 % VOLUME REDUCTION			PERMEATE AT 90 % VOLUME REDUCTION		
			CHP	TFC	PS	CHP	TFC	PS	CHP	TFC	PS
TURB	NTU	1000	2.70	0.73	0.89	1.60	0.56	0.88	3.50	0.80	NA
TOC	mg/L	800	83.0	32.5	64.0	85.5	35.5	77.0	172.0	69.0	127.0
ALKALINITY	mg/L	NA	NA	NA	NA	NA	NA	NA	240	190	237

NO EMULSION BREAKING WAS OBSERVED EVEN AT 90% VOLUME REDUCTION

NA - data not available

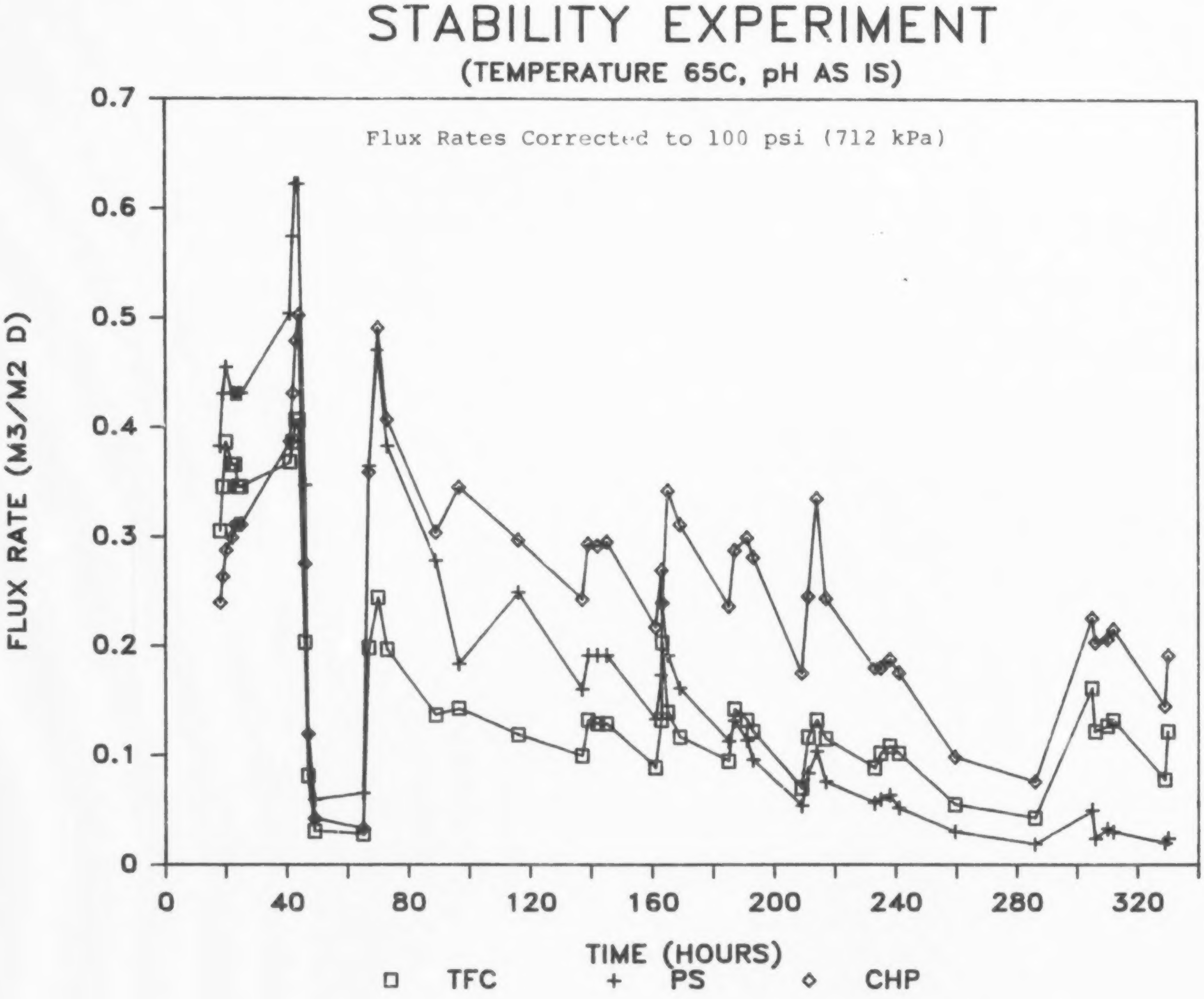
in the permeates of CHP, TFC, and PS were 172 mg/L, 69 mg/L, and 127 mg/L respectively. The concentration of TOC in the permeates of all membranes increased as the feed became more concentrated.

3.5 Long Term Stability Test

The flux rate data for the long term stability test are shown in Figure 3-16. Results from analyses of samples taken during the run are presented in Table 3-15. Complete results are included in Appendix E.

This run was conducted as a continuation of the batch concentration run. A volume reduction of 90% was achieved after 47 hours of operation and from then on the permeates were recycled back to the process tank. The flux rates during the period between 47 and 70 hours were very low. The temperature and flow to the cells during this period of time were both lower than normal. The temperature during this period dropped from 65°C to 55°C and the flow to the cells dropped from 2 L/min to 1.2 L/min. When the temperature and flow to the cells were increased once again to their target values of 65°C and 2 L/min the flux rates for all three membranes increased as well. The flux rates for CHP, PS, and TFC reached values of 0.49, 0.47, and 0.24 m^3/m^2 D respectively. During the following 260 hours the CHP membrane consistently had the highest flux rate and exhibited only a moderate decline with time. The PS membrane started out almost as high as the CHP but exhibited a steeper flux rate decline over the period of the run. The TFC membrane had a lower flux at the start of the 90% reduction stability run than the CHP or PS membranes but had a very stable flux rate. By the end of the run the flux rate for PS had

FIGURE 3 - 16



STABILITY EXPERIMENT

(TEMPERATURE 65C, pH AS IS)

TABLE 3 - 15

LONG TERM STABILITY RUN ANALYTICAL DATA

PARAMETER	UNITS		CONCENTRATE			PERMEATES									ONCE THROUGH BOILER FEED QUALITY
						ONE DAY			ONE WEEK			TWO WEEK			
		FEED	1 DAY	1WEEK	2WEEK	CHP	TFC	PS	CHP	TFC	PS	CHP	TFC	PS	
pH		7.97	8.83	9.01	8.95	8.28	8.25	8.45	8.8	8.65	8.76	8.58	8.55	8.6	NS
COND	uHmos/cm	930	1100	3500	3800	750	440	620	2100	1750	2300	2500	2400	3250	NS
TURB	NTU	1000	7300	44000	17000	7.2	0.6	1.2	2.0	3.4	2.0	2.8	3.1	NA	NS
TOC	mg/L	880	1370	10100	7040	83	41	78	166	137	280	178	163	280	NS
TDS	mg/L	1117	1562	19964	11440	728	462	606	2516	1592	2640	1860	1300	2580	< 8,000
TSO	mg/L	1127	3222	6674	1517	NA	NA	NA	NA	NA	NA	NA	NA	NA	< 1.0
TSS	mg/L	44	125	1720	425	NA	NA	NA	NA	NA	NA	NA	NA	NA	0.0
DIS SIO2	mg/L	84	112	NA	NA	69.4	61.5	64.0	62.2	67.2	73.2	32.7	40.0	40.2	< 50-150
HARDNESS	mg/L	12.0	72.0	211.0	133.0	116.0	0.4	0.6	3.6	3.3	6.0	2.9	1.8	7.4	< 0.5

NS - not specified
NA - analysis not done (TSO & TSS essencially zero for permeates)

decreased to be significantly less than that for the TFC membrane. Least square analysis of the flux rate data from 80 hours to the end of the run gives flux rate declines for CHP, PS, and TFC of 0.016, 0.024, and 0.003 m^3/m^2 D per day respectively. The high and relatively stable flux for the CHP membrane makes it an attractive membrane for further investigation.

As expected the analytical data for the long term stability runs show the best rejections for all components by the TFC membrane and similar rejections for the CHP and PS membranes. There appears also to be no significant decrease in rejection over time for any of the membranes.

The analysis of the concentrated emulsion in the process tank after one week and after two weeks shows a significant decrease in almost all components with time. Suspended oil and suspended solids decreased 77% and 75% respectively during the second week of operation. The concentration of all components except conductivity decreased. It was observed after the run that significant quantities of bitumen had coated most of the piping in the system and the pump internals. Since the total volume of concentrated emulsion was low, only about 4 liters, it is possible that oil and suspended solids were being lost to the inner surface of the system piping. The decrease in TSO during the second week of operation can be explained by only 20 grams of bitumen sticking to the piping.

During the run several attempts were made to clean the membranes by increasing the flow to the cells for short periods of time. After such flushing, the flux rates for all membranes increased slightly but only for a short period of time.

3.6 **Reverse Osmosis Tests**

3.6.1 RO Screening Test

Results of the RO membrane screening test using PS-UF membrane permeate as a feed are shown in Figure 3-17. Appendix F contains complete data for this test including data from the pre and post salt water flux test.

During the 4 1/2 hour initial salt water test the flux rates for the thin film composite (TFC) and cellulose acetate (CA) membranes decreased from 1.76 to 0.96 and from 1.41 to 0.69 m^3/m^2 D respectively. The rejections of conductivity for TFC and CA membranes were 97.4% and 96.3%.

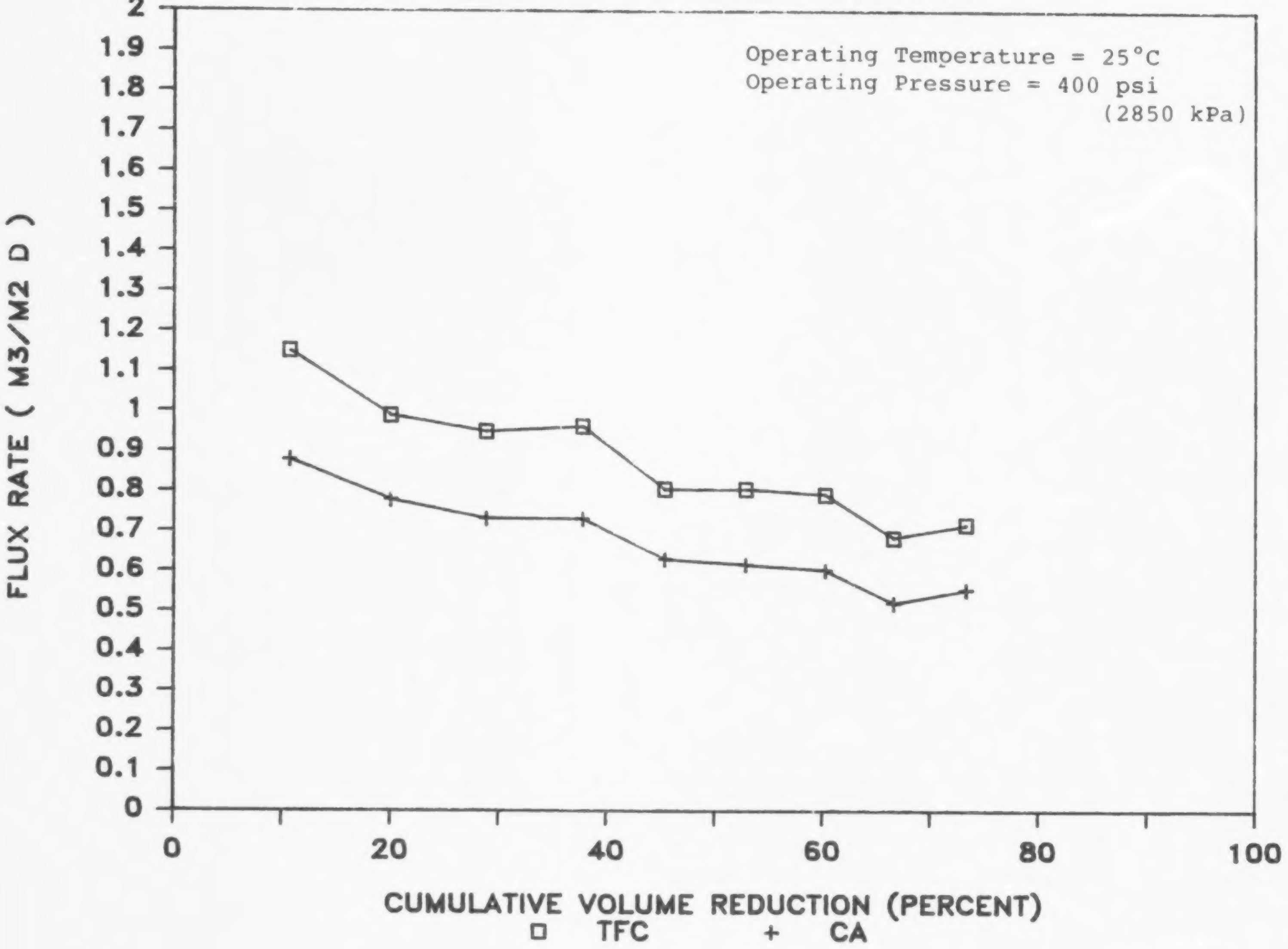
The system was then operated with UF permeate from a polysulphone (PS) membrane which had been operated with a feed of bitumen emulsion. During the 4 1/2 hours of this test the flux rates for TFC and CA membranes decreased from 1.15 to 0.72 and from 0.88 to 0.555 m^3/m^2 D respectively. The volume of feed was reduced by 73% during this time. The rejections of conductivity for the TFC and CA membranes were 98.6% and 97.8%. Also during the run the pH of the feed being concentrated increased from 8.2 to 8.5 and the pH of both the permeates decreased over the period of the run from about 7.1 to 6.5.

The flux rates for both membranes were relatively stable during the final salt water flux test. Flux rates during the 4 1/2 hour test for TFC and CA membranes went from 0.93 to 1.04 and from 0.755 to 0.69 m^3/m^2 D respectively. Rejections of conductivity were 97.7% and 94.9% for TFC and CA.

FIGURE 3 - 17

RO MEMBRANE SELECTION

FLUX RATE VS VOLUME REDUCTION



The TFC membrane had flux rates and conductivity rejections consistently better than the CA membrane. The decreases in flux during the initial salt tests are believed to be primarily the result of membrane compaction. The decrease in flux during the UF permeate run is believed to be the result of membrane compaction and a steadly increasing concentration of components in the feed.

Based on the results of this test the TFC membrane was chosen for use in further studies on RO treatment for final polishing of produced water.

3.6.2 Reverse Osmosis Test on Ultrafiltration Permeate

Flux rate data for the test of the TFC-RO membrane using TFC-UF membrane permeate as a feed are shown in Figure 3-18. Table 3-16 shows the results of analysis of the feed and of permeate samples taken during the test. Complete test results including salt water flux test data are presented in Appendix F.

The flux rate for the two hour initial salt water flux test started at 1.38 m^3/m^2 D, rose to 2.27 m^3/m^2 D after 45 minutes, then decreased to 1.22 m^3/m^2 D by the end of the run. The mean conductivity rejection during the test was 97.6%.

The RO system was then fed with permeate generated during a run using a UF TFC membrane, G10, with a feed of bitumen emulsion. During this run the flux rate through the RO TFC membrane was relatively steady at 1.35 m^3/m^2 D until the very end of the run when the flux rate dropped to 0.99 m^3/m^2 D. The volume reduction at the end of the run was 91% (concentration factor 11.1). The average rejection of conductivity during the run was 99.0%. The pH of the

FIGURE 3 - 18

RO TREATMENT OF UF PERMEATE

TFC FLUX RATE VS VOLUME REDUCTION

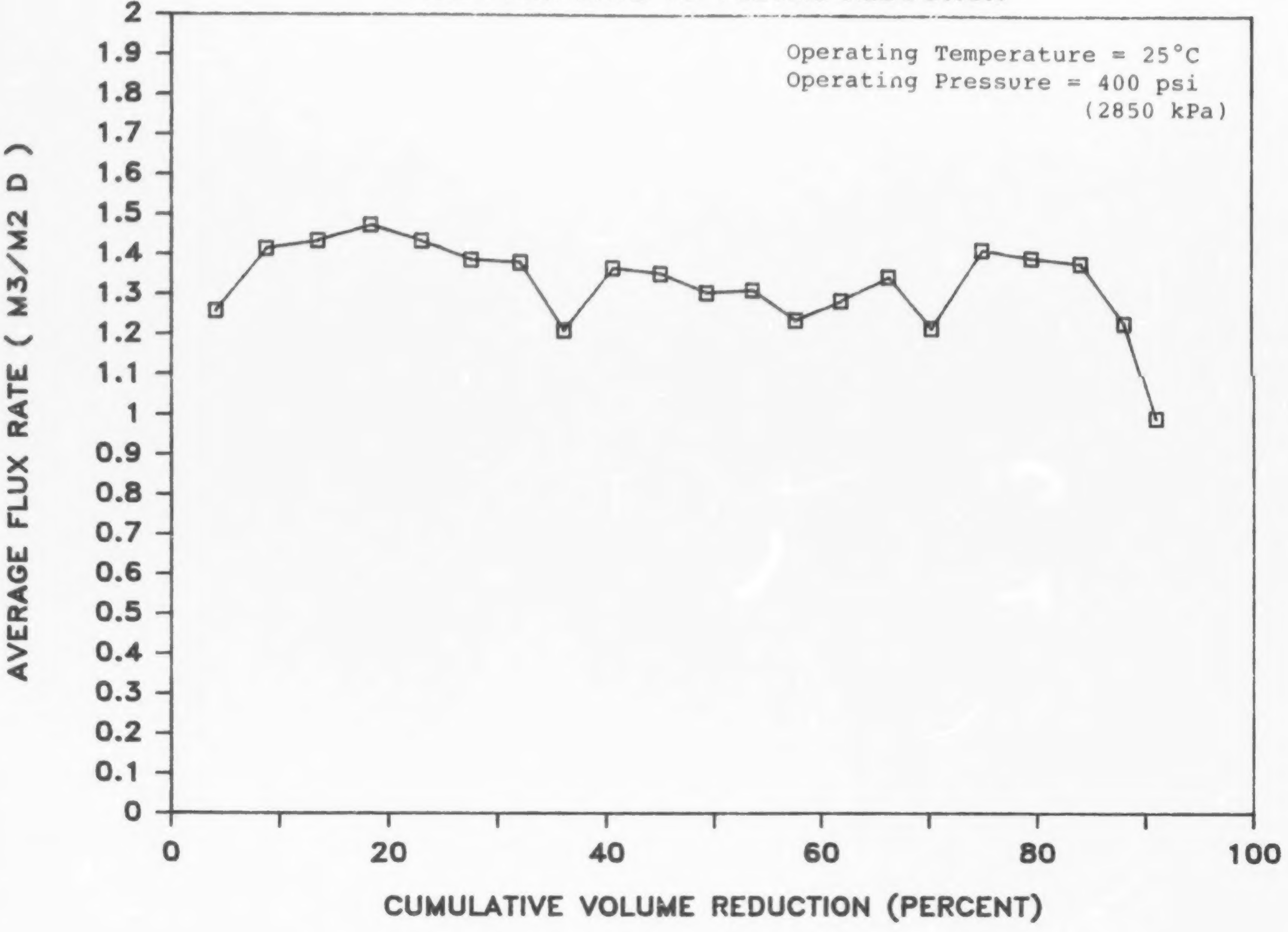


TABLE 3 - 16

ANALYTICAL DATA FOR RO TREATMENT OF UF PERMEATE

PARAMETER	UNITS	CONCENTRATE		PERMEATE			
		FEED	FINAL	30 MIN	30%	60%	91%
pH		8.55	8.60	6.00	6.17	6.08	6.60
COND	uHmos/cm	1100	2900	13	12	13	27
TURB	NTU	0.75	6.00	0.05	0.07	0.09	0.35
TOC	mg/L	75	122	10.2	4.5	1.5	1
TDS	mg/L	880	2630	<1	<1	20	30
HARDNESS	mg/L	2.82	16.78	0.79	0.83	0.50	0.29
DIS SIO2	mg/L	63.03	120.87	1.60	1.74	1.95	3.85
F	mg/L	5.20	17.80	0.03	0.03	0.04	0.09
Cl	mg/L	260.0	660.0	1.74	0.87	1.82	5.20
NO2	mg/L	<1	<1	<.1	<.1	<.1	<.4
PO4	mg/L	<0.1	<0.1	<.01	<.01	<.01	<.01
NO3	mg/L	<.04	<.04	<.04	<.04	<.04	<.04
SO4	mg/L	36.00	76.00	0.09	0.15	0.18	0.13
Be	mg/L	<.005	<.005	<.005	<.005	<.005	<.005
Mo	mg/L	0.020	0.070	<.005	<.005	<.005	<.005
Ca	mg/L	0.90	5.30	0.20	0.25	0.15	0.10
V	mg/L	<.005	0.015	<.005	<.005	<.005	<.005
Al	mg/L	0.05	0.17	<.01	0.01	<.01	0.01
Mg	mg/L	0.14	0.86	0.07	0.05	0.03	0.01
Ba	mg/L	<.005	0.015	0.005	<.005	<.005	<.005
K	mg/L	7.80	24.00	0.04	0.10	0.14	0.16
Sr	mg/L	0.005	0.020	<.005	<.005	<.005	<.005
Na	mg/L	280.0	800.0	2.4	1.7	2.3	6.1
Zn	mg/L	<.005	0.085	0.030	0.020	0.055	0.010
Cd	mg/L	<.005	0.045	0.050	0.050	0.050	0.030
Mn	mg/L	<.005	0.005	0.005	<.005	0.015	<.005
Co	mg/L	<.005	0.005	<.005	<.005	<.005	0.005
Cu	mg/L	0.020	0.090	0.010	<.005	<.005	<.005
Ag	mg/L	<.01	<.01	<.01	<.01	<.01	<.01
Fe	mg/L	0.06	1.44	0.04	0.02	0.04	0.05
Pb	mg/L	0.02	0.04	<.01	<.01	<.01	0.01
Cr	mg/L	<.005	0.110	<.005	<.005	<.005	<.005
Ni	mg/L	<.005	0.105	<.005	0.005	<.005	0.010

Operating Temperature = 25 deg. C
Operating Pressure = 400 psi (2850 KPa)

feed increased from 7.8 to 8.6 as it became more concentrated. The pH of the permeates remained close to 7.0. The detailed analysis of the permate showed that a very high quality permeate is generated by the TFC RO membrane. The concentrations of TOC and hardness in the permeate at a 30% volume reduction were 4.5 mg/L and 0.83 mg/L. The concentration of these components in the permeate at a 91% volume reduction decreased to 1 mg/L and 0.29 mg/L. The concentrations of other components in the permeates increased as they became more concentrated in the feed. The permeate at 91% volume reduction does, however, meet the requirements specified for the once through boiler systems. Conductivity, TDS, and dissolved silica in the permeates at 91% volume reduction were 27 microHmos/cm, 30 mg/L, and 2.85 mg/L respectively.

Infrared analysis of the RO permeate indicated a strong O-H stretch at 3750-3000 cm^{-1} but little evidence of a carbonyl stretch at 1700-1740 cm^{-1} or stretches at 1450, 1410 or 1370 associated with organic acids. Gas chromagraph analysis showed little sign of any organics in the sample.

The final salt water flux test showed a lower flux (0.87 m^3/m^2 D) than was observed during the initial salt test (2.27 - 1.22 m^3/m^2 D). The rejection of conductivity during this test was 97.4%. The flux rate during the test on UF permeate showed a sudden decrease after an 85% volume reduction (concentration factor of 6.67) had been obtained. The solubility of silica at 25°C at a pH of 8.6 is approximately 160 mg/L. It is believed that the decrease in flux at the end of the UF permeate test and the low salt water flux in the final test is the result of silica precipitation plugging membrane pores.

The problems with silica precipitation can be decreased by increasing the operating temperature and increasing the pH of the feed. At a temperature of 40°C and a pH of 10 the solubility of silica is increased to over 560 mg/L. Figure 3-19 shows silica solubilities as a function of temperature and pH. To avoid precipitation of silica it is suggested that the concentration in the concentrate be limited to 80% of the maximum solubility. Operation at 40°C at pH 10 should allow concentration of the feed up to a concentration factor of seven (560 x 0.8 ÷ 63 = 7.1) or a percent volume reduction of 86% without problems of silica precipitation.

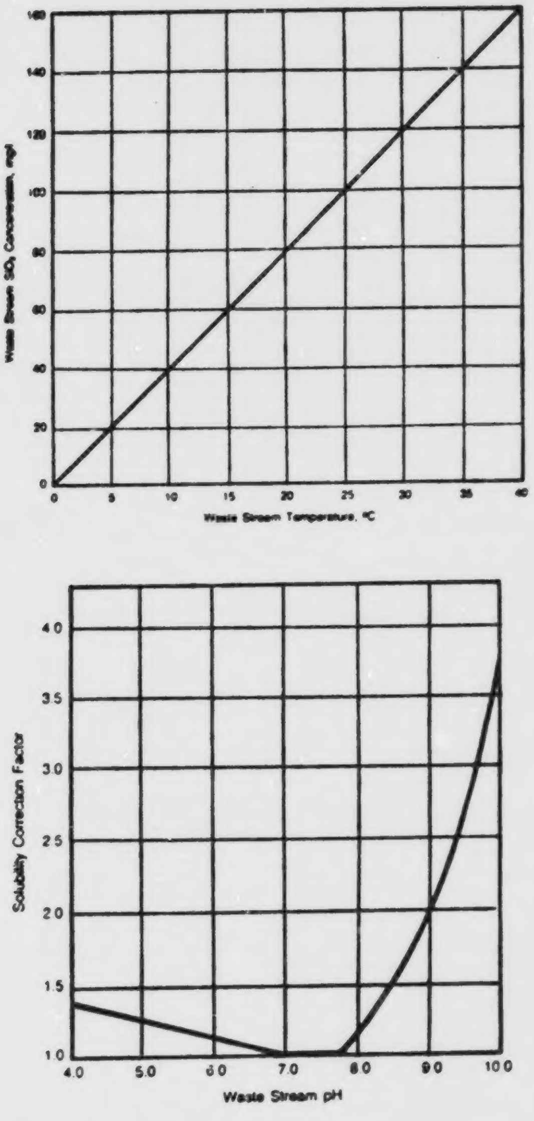
Barium and strontium sulfates may also be a source of scaling in RO systems. Special precautions to prevent scaling may be required when the feed concentrations of Ba or Sr are greater than 0.1 mg/L. The concentrations of these elements in the UF permeate and demulsified water are less than 0.1 mg/L but careful monitoring of different water sources should be conducted to insure that such scaling does not occur.

3.6.3 Reverse Osmosis Test on Demulsified Bitumen Emulsion

Flux rate data for the test of the TFC-RO membrane using demulsified filtered produced water as a feed are shown in Figure 3-20. Table 3-17 shows the results of analyses of the feed and of permeate samples taken during the run. Complete results including salt water flux test data are presented in Appendix F. The TFC-RO

FIGURE 3 - 19

SILICA SOLUBILITY AS A FUNCTION OF TEMPERATURE AND pH



from Rak (1984)

FIGURE 3 - 20

RO (DEMULSIFIED PRODUCED WATER)

TFC FLUX RATE VS VOLUME REDUCTION

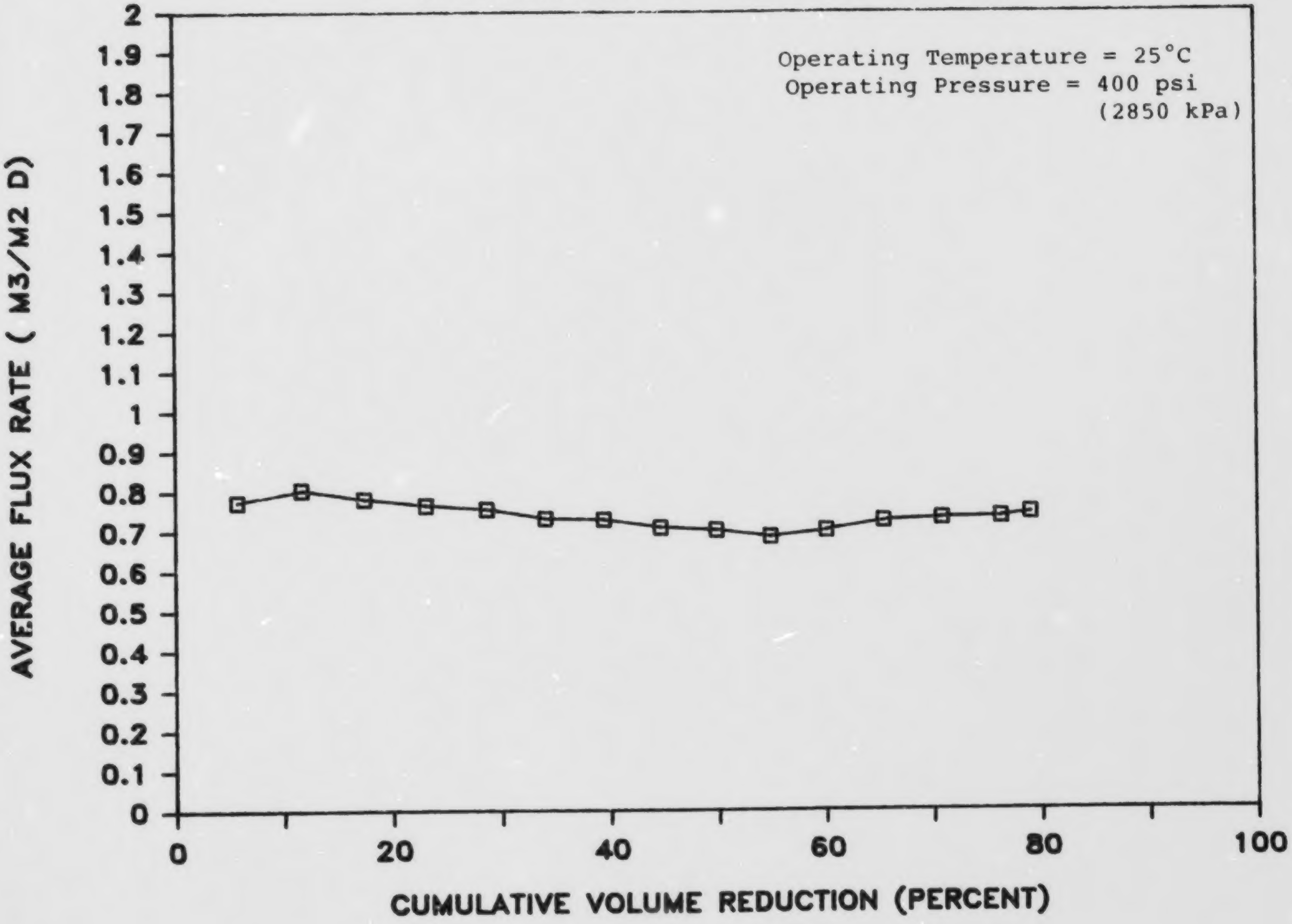


TABLE 3 - 17

ANALYTICAL DATA FOR RO TREATMENT OF DEMULSIFIED PRODUCED WATER

PARAMETER	UNITS	CONCENTRATE FEED	FINAL	30 MIN	PERMEATE 30%	60%	80%
pH		7.72	8.82	5.90	6.50	6.04	6.60
COND	uHmos/cm	980	4800	24	16.4	18.7	43.5
TURB	NTU	6.80	120.00	0.85	0.34	0.24	0.74
TOC	mg/L	135	600	2	3	4	15
TDS	mg/L	720	719	49	34	43	68
HARDNESS	mg/L	2.62	20.75	1.50	1.58	0.75	0.87
DIS SIO2	mg/L	59.60	330.30	0.70	0.70	0.90	1.50
F	mg/L	4.10	28.00	0.07	0.07	0.08	0.08
Cl	mg/L	192.0	1050.0	5.10	1.70	3.10	8.00
NO2	mg/L	<0.1	<5	<0.05	<0.05	<0.05	<0.05
PO4	mg/L	0.30	3.80	0.20	0.30	<0.2	0.20
NO3	mg/L	<0.2	0.20	<0.2	<0.2	<0.2	0.60
SO4	mg/L	78.00	480.00	0.50	0.20	0.20	<.2
Be	mg/L	<0.02	<0.02	<0.005	<0.005	<0.005	0.005
Mo	mg/L	0.040	0.260	<0.005	<0.005	<0.005	<0.005
Ca	mg/L	0.80	6.00	0.50	0.55	0.25	0.35
V	mg/L	<0.02	<0.02	<0.005	<0.005	<0.005	<0.005
Al	mg/L	0.09	0.30	0.05	0.13	0.04	0.09
Mg	mg/L	0.15	1.40	0.06	0.05	0.03	<0.01
Ba	mg/L	0.020	0.080	0.005	<0.005	0.010	<0.005
K	mg/L	6.20	37.00	0.11	0.02	0.05	0.11
Sr	mg/L	<0.02	0.060	<0.005	<0.005	<0.005	<0.005
Na	mg/L	210.0	1160.0	4.4	2.4	3.6	8.0
Zn	mg/L	0.020	0.330	0.030	0.015	0.045	<0.005
Cd	mg/L	<0.01	0.010	0.020	0.010	0.005	<0.005
Mn	mg/L	0.100	0.440	0.005	0.010	0.010	<0.005
Co	mg/L	<0.01	<0.01	<0.005	<0.005	<0.005	<0.005
Cu	mg/L	0.030	0.360	0.010	0.010	0.020	0.010
Ag	mg/L	<0.01	<0.03	<0.01	<0.01	<0.01	<0.01
Fe	mg/L	0.83	2.90	0.10	0.05	0.22	0.03
Pb	mg/L	<0.03	<0.03	0.01	0.02	0.02	0.01
Cr	mg/L	<0.01	0.040	<0.005	<0.005	0.005	<0.005
Ni	mg/L	0.020	0.150	<0.005	0.005	<0.005	0.015

Operating Temperature = 25 deg. C
Operating Pressure = 400 psi (2850 KPa)

membrane used was a test membrane of the same type as was used for the UF permeate test.

Flux rates for the initial salt water flux test were stable and averaged 0.96 m^3/m^2 D. This is less than even the minimum flux observed during the initial salt water flux test for the UF permeate test which was 1.22 m^3/m^2 D. This lower flux may be due to inconsistencies in the membrane sheet or may be the result of a lower flowrate through the test unit. The percent rejection of conductivity during the test was 98.0%.

During the RO test using the demulsified produced water the flux rate remained stable near 0.74 m^3/m^2 D. This low flux rate is thought to be the result of the same factors which resulted in a lower initial salt water flux and the result of the slightly different feed water characteristics of the demulsified produced water. The run was stopped at an 80% volume reduction to avoid silica precipitation problems. The average rejection of conductivity during the run was 99.1%.

The detailed analyses of the feed water demonstrated that demulsified and filtered produced water has similar characteristics to TFC-UF permeate except that more oil is present in the demulsified water as shown by the higher turbidity. The demulsified water had a turbidity of 6.8 NTU, a TOC level of 135 mg/L, and an iron content of 0.83 mg/L. The UF permeate had turbidity of 0.75 NTU, TOC of 75 mg/L, and iron at 0.06 mg/L. The concentrations of other components in the demulsified water were very similar to those present in the UF permeate. The rejections of contaminants in this test were very similar to the rejections obtained for the previous RO test. The permeate does still meet the quality criteria for the once through boiler system.

The flux rates during the final salt water test were stable at about 1.07 m^3/m^2 D, up slightly from the 0.96 m^3/m^2 D observed during the initial test. It is apparent that the silica scaling problem encountered in the previous run was not occurring during this test.

4.0 ECONOMIC ASSESSMENTS

4.1 Basis of Costing

Three separate processes for producing boiler feedwater from produced water have been outlined and estimates of capital and operating costs for each have been prepared. The processes involve:

i) using an ultrafiltration (UF) system followed by a softener,

ii) using a UF system followed by a reverse osmosis (RO) system, and

iii) using conventional technology for oil removal (induced gas floatation (IGF) and filtration) followed by RO.

The capacities of the processes are based on the same capacity system as was used by CH2M HILL in their 1982 Report, "Generation of Steam Using Low Grade Fuels and Field Produced Water for In-Situ Oil Recovery". The once through 80% boiler in the report was rated at 73 MW or 250 M BTU/h and required a feed of approximately 500,000 US gallons per day (USGPD). All processes have been designed to produce 500,000 US GPD of boiler feed water.

The designs of the steps used for pre and post membrane treatment were developed based on information available on present treatment practices from the literature and contacts with oil industry representatives.

The purchase costs for equipment other than the membrane process steps were based on estimates provided by equipment suppliers or from process equipment cost correlation tables. The total costs for process equipment were developed using factors of the purchase costs to estimate the other direct and indirect cost of installed equipment. The installed costs include all costs associated with the facility including foundations, instruments, piping, insulation, buildings, and services, as well as construction and field expenses, engineering, and contractors fees. The costs for membrane systems presented in Appendix G are based on the results of experimental work and ZENON's up-to-date cost base information for membrane systems.

4.1.1 Ultrafiltration/Softening Process

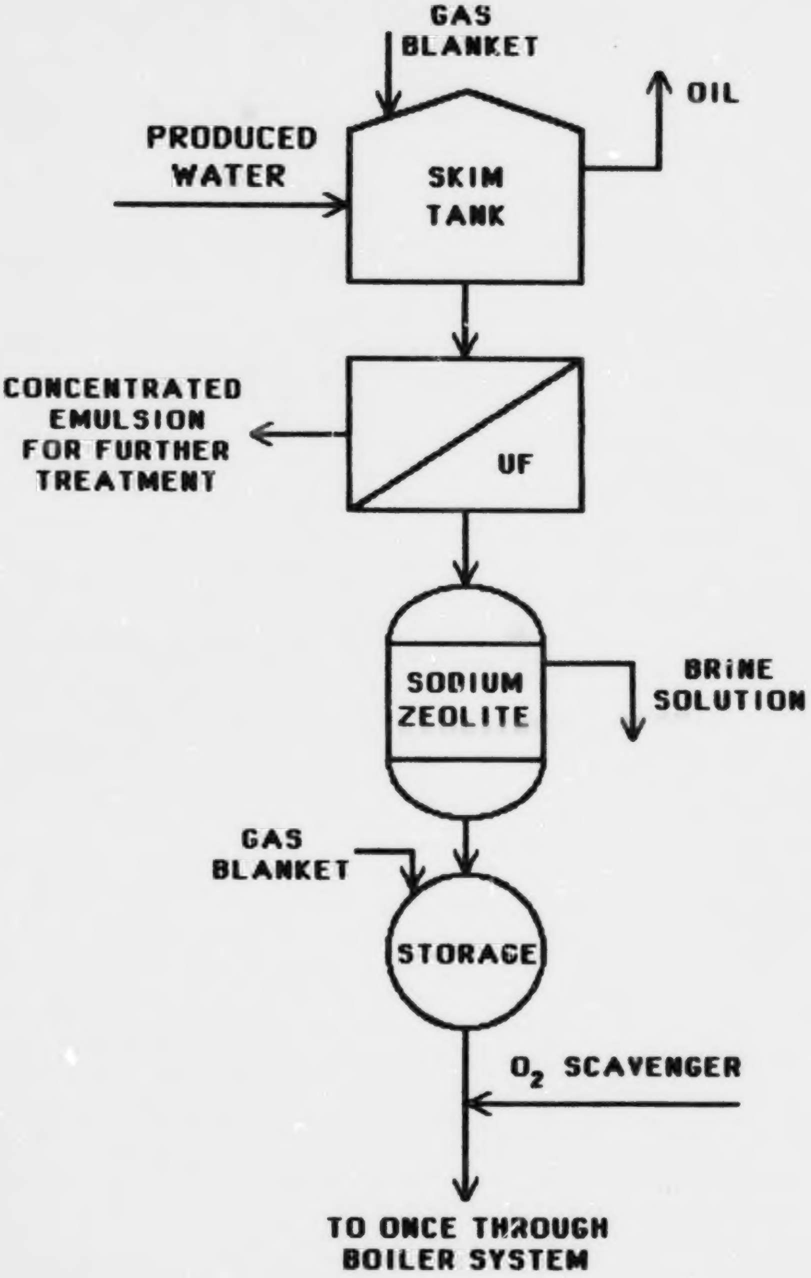
Figure 4-1 shows a flow schematic for the ultrafiltration/softening (UF/SOFT) process for produced water treatment. The skim tank has been designed to have a two hour retention time and is equipped with a surface oil skimmer, a mechanism for bottom sludge removal, and a natural gas blanket. It is anticipated that this step will remove all free and suspended oil and produce an effluent containing approximately 1700 mg/L of emulsified oil.

Two ultrafiltration (UF) systems one using 8" spiral wrap membranes and the other using 1/2 inch tubular membranes were designed based on the results of experimental work and ZENON's experience at designing membrane systems. The design flux rates for the tubular system was estimated at 0.55 m^3/m^2 D. During the batch concentration run a flux of 0.62 m^3/m^2 D was obtained with the PS membrane at a volume reduction of 82% (CF of 5.6). Although the flux rates during the long term stability run at 90% volume reduction (CF of 10) were lower (0.47 m^3/m^2 D), it is reasonable to expect that by concentrating only to an 80% volume reduction (CF of 5) and optimizing the operating conditions a flux of 0.55 m^3/m^2 D can be maintained. ZENON's experience with tubular and spiral systems suggests that the flux

FIGURE 4 - 1

ULTRAFILTRATION/SOFTENING PROCESS

FOR PRODUCED WATER TREATMENT


GAS BLANKET
OIL
PRODUCED WATER
SKIM TANK
CONCENTRATED EMULSION FOR FURTHER TREATMENT
UF
SODIUM ZEOLITE
BRINE SOLUTION
GAS BLANKET
STORAGE
O_2 SCAVENGER
TO ONCE THROUGH BOILER SYSTEM

rates which are obtained for spiral systems are generally 70% of those obtained with tubular systems. This lower flux is a result of somewhat poorer flow characteristics in the spiral systems. The design flux rate for the spiral UF system was 0.385 m^3/m^2 D. Both flux rates are considered to be conservative estimates and are toward the lower end of the range of acceptable flux rates for membrane systems. The system costs are virtually a direct function of the flux rate obtained therefore any increase in flux rate will reduce the system costs considerably.

The UF processes have been designed for 80% recovery of the feed. This recovery rate has been specified in order to maintain higher membrane flux rates and reduce fouling problems associated with very high rates of recovery. Based on the flux rate and permeate flow required detailed costing of the UF systems has been conducted. Appendix G presents a flow schematic and the purchase costs of the membrane systems.

The sodium zeolite softening system specified is a less expensive alternative to multi-step softening required in other produced water treatment processes. Such a system could be used since hardness and TDS levels in the UF permeate are expected to be low. The sodium zeolite system has been designed to reduce hardness levels to less than 0.5 mg/L.

A storage tank with a 24 hour nominal retention time has also been specified. Provisions have also been made for the addition of 50 mg/L of sodium sulfite as an oxygen scavenger.

4.1.2 Ultrafiltration/Reverse Osmosis Process

Figure 4-2 shows a flow schematic for the UF/RO process for produced water treatment. The skim tank and UF systems are identical to those in the process using UF/softening except that the UF system has a higher capacity to account for the loss of water in the RO concentrate.

The RO system was designed based on flux data generated during the experimental work. A spiral RO system was costed in detail using 8" diameter spiral modules with an 80% recovery rate to insure silica precipitation does not occur. The assumed flux rate based on the flux obtained at an 80% volume reduction during the RO test was 1 m^3/m^2 D.

Storage and oxygen scavenging were also specified in this flow sheet.

4.1.3 Induced Gas Floatation/Filtration/Reverse Osmosis Process

Figure 4.3 shows a flow schematic for the IGF/RO process for produced water treatment. The skim tank is designed based on a two hour retention time and as before is equipped with a surface skim and bottom sludge removal. Chemical demulsifier WITCO RTC 323 at 200 mg/L is added to the feed of the skim tank resulting in a much lower skim tank effluent oil concentration than with the previous systems. The skim tank is followed by a four chamber induced gas floatation unit, an intermediate storage tank, a sand filter, and cartridge filter. The cartridge filter is in place to protect the RO membrane from oil which may get through the sand filter during upset conditions.

FIGURE 4 - 2
ULTRAFILTRATION/REVERSE OSMOSIS PROCESS
FOR PRODUCED WATER TREATMENT

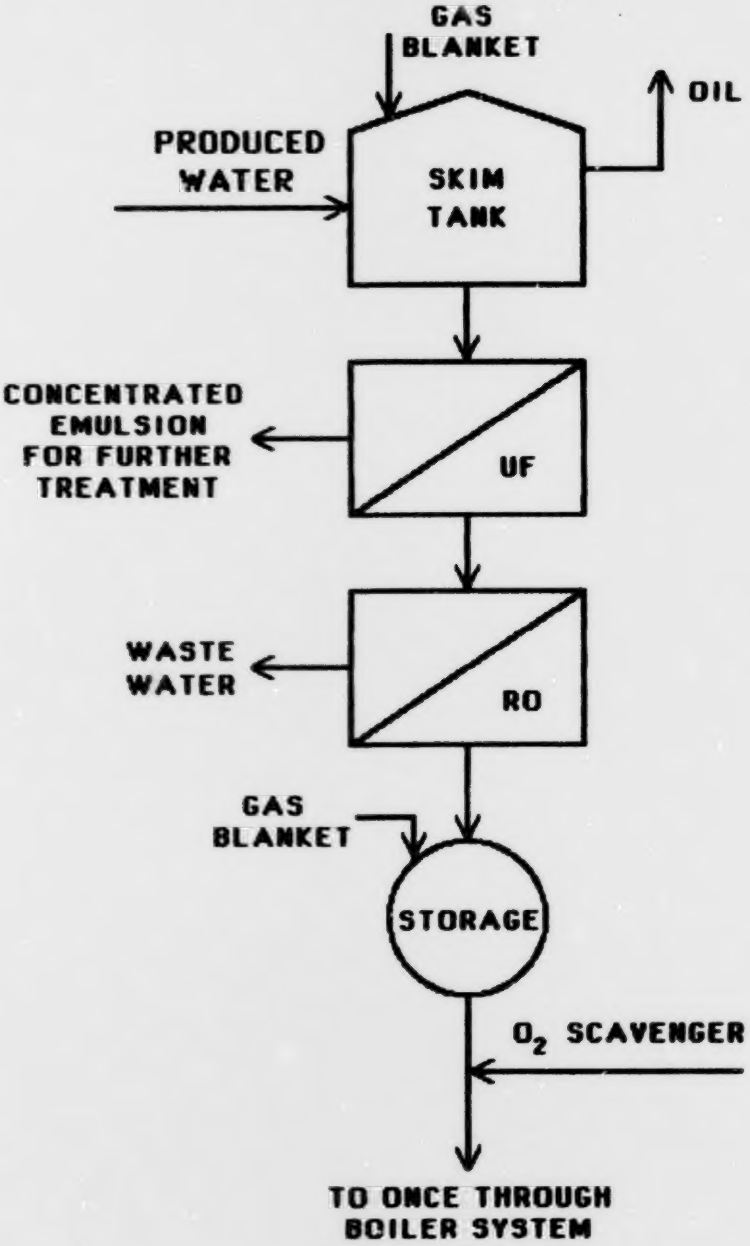
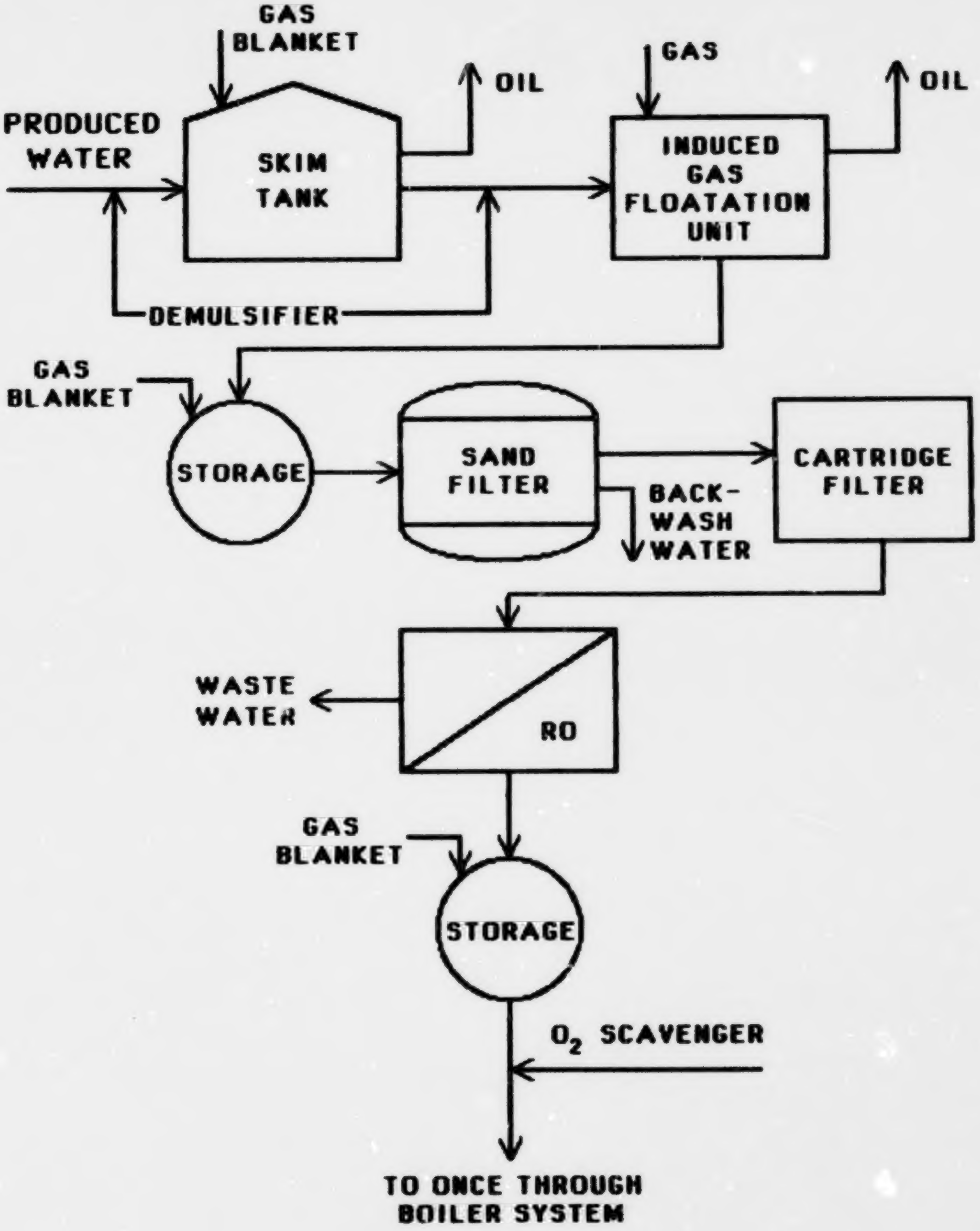

GAS BLANKET
OIL
PRODUCED WATER
SKIM TANK
CONCENTRATED EMULSION FOR FURTHER TREATMENT
UF
WASTE WATER
RO
GAS BLANKET
STORAGE
O_2 SCAVENGER
TO ONCE THROUGH BOILER SYSTEM

FIGURE 4 - 3

INDUCED GAS FLOATATION/FILTRATION/REVERSE OSMOSIS PROCESS
FOR PRODUCED WATER TREATMENT


GAS BLANKET
OIL
GAS
OIL
PRODUCED WATER
SKIM TANK
INDUCED GAS FLOATATION UNIT
DEMULSIFIER
GAS BLANKET
STORAGE
SAND FILTER
CARTRIDGE FILTER
BACK-WASH WATER
WASTE WATER
RO
GAS BLANKET
STORAGE
O_2 SCAVENGER
TO ONCE THROUGH BOILER SYSTEM

The RO unit has been designed using 8" spiral membranes based on a flux rate of 0.775 m^3/m^2 D and 80% recovery of the feed. A tubular RO system was considered for this application but was felt to be both unnecessary and too expensive in comparison to the spiral membrane system.

Final storage and oxygen scavenging were also specified for this system.

4.2 Cost Estimates

Appendix G shows the details of the capital and operating costs for each step in the three processes for produced water treatment. Tables 4-1, 4-2, and 4-3 show summaries of these results for each of the three processes.

The total costs including amortization of capital at 13% over 10 years are very similar for the spiral UF/RO and IGF/RO processes. The costs for both processes are close to \$7.00 per thousand gallons treated. The cost for the spiral UF/Softening process is only \$4.84 per thousand gallons treated.

The sodium zeolite softening process specified in the UF/SOFT process is inexpensive relative to the softening required after conventional treatment of typical produced waters. The low hardness in the feed and reduction of TDS with the UF membrane allows for the much simpler sodium zeolite process to be used. The processes which had RO polishing are more expensive than the ones using only sodium zeolite softening but the quality of the effluent from the RO is much higher. The effluent from the RO exceeds the minimum criteria for the once through boiler feedwater quality considerably. This

TABLE 4 - 1

Total Costs for Ultrafiltration Process for Produced Water Treatment

Capital Costs

	Spiral UF	Tubular UF
Pretreatment	$535,184	$535,184
UF System	$2,061,640	$8,888,799
Softening System	$205,120	$205,120
Storage & Oxygen Removal	$249,652	$249,652
Total Capital Costs	$3,051,596	$9,878,755
Amortized Capital Costs (over 10 years at 13% interest)	$561,494	$1,817,691

Operating and Maintenance Costs (per year)

	Spiral UF	Tubular UF
Pretreatment	$2,000	$2,000
UF System	$200,443	$692,594
Softening System	$25,475	$25,475
Storage & Oxygen Removal	$94,275	$94,275
Total Annual O & M Costs	$322,193	$814,344

Total Costs Including Capital Amortization

	Spiral UF	Tubular UF
Annual Costs	$883,687	$2,632,035
Cost Per Thousand Gallons Treated	$4.84	$14.42

TABLE 4 - 2

Total Costs for Ultrafiltration / Reverse Osmosis Process for Produced Water Treatment

Capital Costs

	Spiral UF	Tubular UF
Pretreatment	$535,184	$535,184
UF System	$2,612,996	$11,040,414
RO System	$995,918	$995,918
Storage & Oxygen Removal	$249,652	$249,652
Total Capital Costs	$4,393,750	$12,821,168
Amortized Capital Costs (over 10 years at 13% interest)	$808,450	$2,359,095

Operating and Maintenance Costs (per year)

	Spiral UF	Tubular UF
Pretreatment	$2,000	$2,000
UF System	$248,698	$880,472
RO System	$115,515	$115,515
Storage & Oxygen Removal	$94,275	$94,275
Total Annual O & M Costs	$460,488	$1,092,262

Total Costs Including Capital Amortization

	Spiral UF	Tubular UF
Annual Costs	$1,268,938	$3,451,357
Cost Per Thousand Gallons Treated	$6.95	$18.91

TABLE 4 - 3

Total Costs for Induced Gas Floatation / Reverse Osmosis Process for Produced Water Treatment

Capital Costs	
Pretreatment	$1,464,897
RO System	$1,112,777
Storage & Oxygen Removal	$249,652
Total Capital Costs	$2,827,326
Amortized Capital Costs (over 10 years at 13% interest)	$520,228
Operating and Maintenance Costs (per year)	
Pretreatment	$551,230
RO System	$123,523
Storage & Oxygen Removal	$94,275
Total Annual O & M Costs	$769,028
Total Costs Including Capital Amortization	
Annual Costs	$1,289,256
Cost Per Thousand Gallons Treated	$7.06

higher than minimum quality feedwater may have an effect in reducing operating and maintenance costs in the boiler. The RO effluent is so clean in fact that it may also be worth considering further deionization treatment and use in utility boilers. The RO treatment also has the ability to remove higher levels of hardness present in many produced waters.

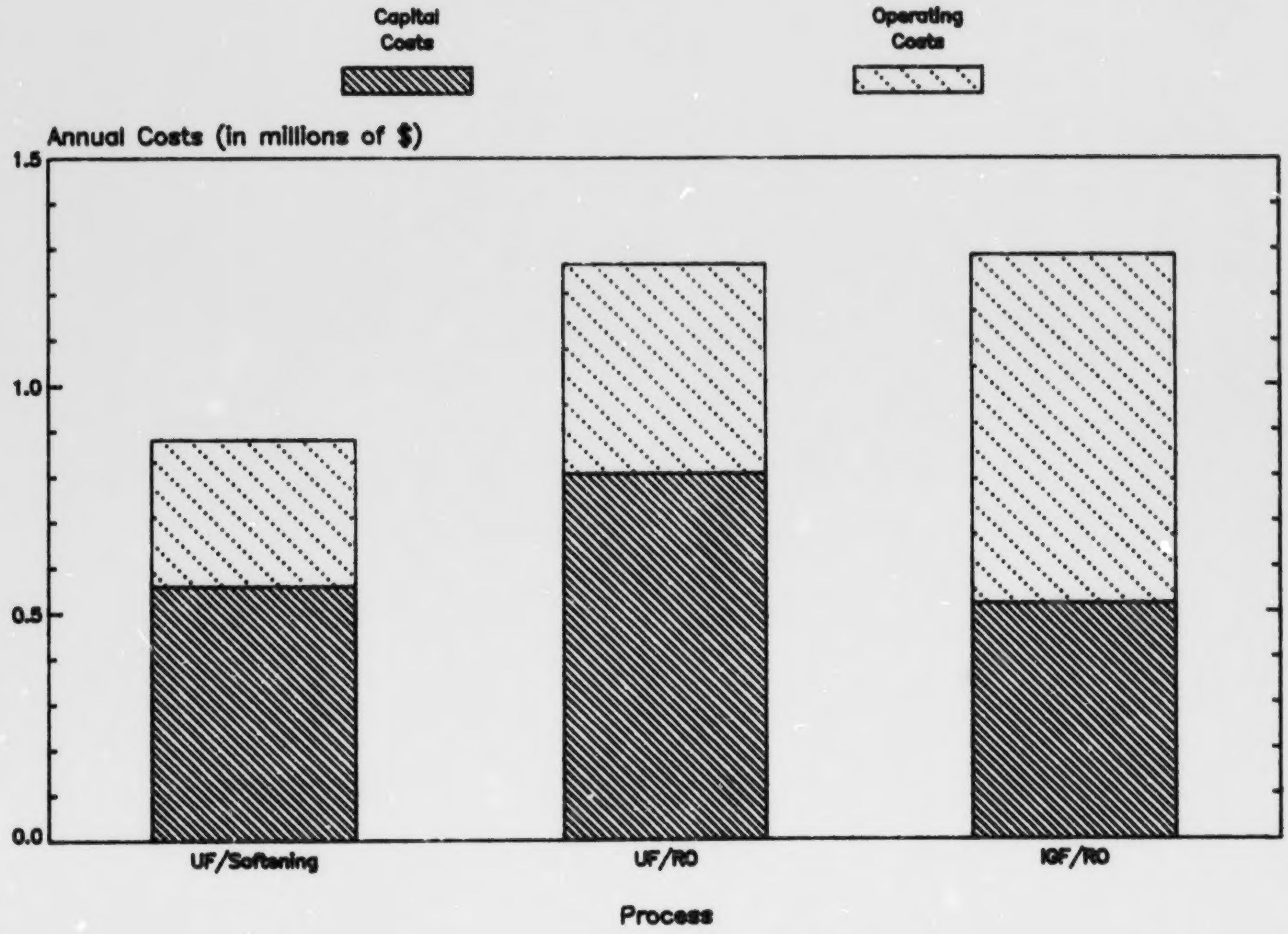
Figure 4-4 shows the yearly amortized capital and operating costs for the three processes. Spiral UF systems have been used in the two processes employing UF. The total costs for the spiral UF/RO and IGF/RO are similar but the UF/RO has a larger proportion (64%) of capital cost than the IGF/RO processes (40%). The UF membrane process has a higher capital cost but much lower operating costs than conventional IGF treatment. The UF/SOFT process is less expensive overall but the quality of the effluent is lower than from the RO processes.

If the cost of the RO pretreatment by IGF and filtration is combined with the sodium zeolite softening the total capital cost is $1,919,669 and the operating costs are $670,980 to give a cost per thousand gallons treated of $5.61. This is more expensive than the UF/SOFT process but less than for the UF/RO and IFG/RO processes. The comparison demonstrates that the spiral UF process can be more economical than conventional treatment processes.

In their 1982 report, CH2M Hill specified a skim tank with demulsifier addition followed by IGF for produced water treatment prior to a thermosludge steam injection process. The capital and operating costs for this two stage process were $880,000 and $300,000/year in 1982. The capital and operating cost for the skim

FIGURE 4 - 4

Amortized Capital and Operating Costs
for Different Processes for Treating Produced Water



tank with demulsifier addition followed by IGF, sand filtration, and a cartridge filter estimated in this work are $1,465,000 and $551,230/year respectively. The differences in costs represent the additional cost of intermediate storage and sand and cartridge filtration.

The CH2M Hill report also estimated the capital and operating costs for a complete treatment process to produce once through boiler feed from produced water at $6,880,000 and $985,000/year respectively. If this capital cost in amortized over 10 years at 13% interest, and added to the operating costs, the cost per thousand gallons treated becomes $12.33. This is higher than the estimates for complete systems developed in this study, but the CH2M Hill study assumed a feedwater with a considerably higher level of hardness (over 130 mg/L) and TDS (800 - 20,000 mg/L). This harder water requires a hot lime softening step and a multi-stage ion exchange for deionization of the water. These operations add considerable expense to the overall process. RO system costs for treatment of very hard water could be higher than for the RO systems considered in this study. Extra expense might be required to prevent scaling of Ca and Mg with some very hard waters.

The comparison with CH2M Hill does, however, suggest the costs of treatment using membranes is competative with conventional processes.

The costs of the tubular systems are very high relative to the spiral systems but the costs are a direct function of the flux rate which can be maintained in the system. Flux data used for estimating the size of systems required were based on flux rates in

NRC flat sheet test cells. The flux rates in actual tubular systems may be considerably higher than obtained in the test cells. Tubular systems also have the advantage that mechanical cleaning of membrane surfaces can be carried out. This cleaning may be required to maintain higher more economical flux rates.

5.0 CONCLUSIONS AND RECOMMENDATIONS FOR FURTHER WORK

5.1 Conclusions

- The technical feasibility of using UF membranes for removing bitumen from a produced water emulsion has been demonstrated.
- Emulsion characterization work demonstrated the considerable variability in produced water and the very stable nature of the emulsion. The sample obtained had oil levels typical of higher oil produced waters (1700 mg/L) and an unusually low hardness (4 mg/L). Detailed chemical analysis showed the presence of about 60 mg/L of naphthenic acid which is though to be an important factor in maintaining emulsion stability.
- Screening tests identified three membranes as being most appropriate for this application. The thin film composite (TFC) UF membrane G10 consistently produced the highest quality permeate and maintained a very stable flux rate. The CHP and polysulfone (PS) E500 membranes produced similar quality permeates and maintained high and reasonably stable flux rates.
- As expected during the screening tests the flux rates for all membranes improved as the temperature of operation was increased. There also appeared to be a trend for increasing flux rates at higher pH.
- During the batch concentration test it was demonstrated that flux rates can be relatively stable even at a 5 fold concentration of the raw emulsion.

- The concentrated emulsion remained stable at a 90% volume reduction or at a concentration factor of 10.
- The long term stability run demonstrated that there is no deterioration in permeate quality with time for any of the membranes tested. This run also demonstrated that effective membrane cleaning will be critical to the economic viability of the process.
- The reverse osmosis tests demonstrated that RO treatment is a simple and effective method of producing boiler feedwater from produced water from which oil has been removed. Silica solubitity problems limit the possible recovery to 80-85% for a 60 mg/L dissolved silica feed.
- Based on the estimated flux rates predicted from experimental data the economics of spiral UF membrane systems are competative with conventional treatment processes. Tubular systems will be uneconomical unless the flux rates for such systems can be increased.
- Simple sodium zeolite softening for treatment of low hardness feeds is less expensive than reverse osmosis but the RO is capable of handling higher hardness feeds and producing higher quality effluents.

5.2 Recommendations

The membrane configurations used during all testing to date have been flat sheet NRC test cells or a large plate and frame test unit. If a commercial scale membrane system is to be put in place it will require the use of configured membrane units probably tubular, spiral wrap, or hollow fiber. The flow characterics in these systems will have a major impact on the flux rates which can be maintained and the efficiency of membrane cleaning. Work should be conducted using a variety of membrane configurations similar to individual modules of large systems but on a small scale. The cost of the tubular system described in this work was considerably higher than that of the spiral system but work with small scale tubular systems may show significantly higher fluxes and better cleanability than other systems. Since large systems usually consist of many modules it is possible to test a single module unit and generate data which can be used to accurately predict the performance of large systems. Such testing should be conducted to generate accurate data on flux rates, membrane fouling, and cleaning efficiency for predicting the performance of full scale systems. Economic assessments of membrane systems can be more accurately refined based on this data.

The choice of membranes from the prescreening tests were made based on flux rates, clarity of permeate, and rejection of TOC. In this application it may in fact be desirable to have a very loose membrane with poor rejection of TOC in order that the surfactant material stabilizing the emulsion passes through the membrane and allows the emulsion to break spontaneously. The emulsion could be concentrated using such a loose membrane to see if spontaneous emulsion breaking will take place.

In addition the effects of produced water aging on membrane system operation should be considered. Once laboratory work on small scale configured systems has demonstrated that stable high flux rates can be obtained and that effective cleaning procedures can be developed, the small scale configured systems should be tested at field locations using fresh emulsions.

Further work should also include exploration of the upper temperature limits of membrane system operation to determine exactly what temperatures can be used and what benefits are achieved by this higher temperature operation.

In this work only one sample of emulsion with an oil level of 1700 mg/L from Texaco's Fort McMurry operation was tested. It is known that there is considerable variability in the emulsions from a single source and even more variability between different sources. As well there are a number different stages of conventional treatment after which membrane systems may be applied. Potential sources of water suitable for UF treatment which should be considered are, skim tank effluent (1000-2000 ppm oil), high temperature separator (HTS) effluent (100-500 ppm oil), and induced gas floatation (IGF) effluent (10-40 ppm oil). Samples of these emulsions should be obtained from different sources to assess the general application of the technology. Flux rates for lower oil content emulsions may be higher than for the emulsion tested in this work resulting in lower capital and operating costs for these membranes systems. Potential sources of water suitable for RO treatment include UF permeates and filtered IGF effluents.

6.0 REFERENCES

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TECHNICAL AND ECONOMIC ASSESSMENT OF THE APPLICATION OF MEMBRANE TECHNOLOGY FOR TREATING OIL/WATER/MINERAL EMULSIONS

FINAL REPORT

VOLUME III - APPENDICES

Report Prepared for

CANMET/ERL

Department of Energy, Mines, and Resources, Canada

Contract No. 0SQ83-00203

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ZENON ENVIRONMENTAL INC.

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ANALYTICAL PROCEDURES

APPENDIX A

ANALYTICAL PROCEDURES

pH

The pH was determined using an Extech Model 651 digital meter (Extech, Boston, Mass.) following the method described in the operating manual.

Conductivity

The conductivity was determined using a Beckman conductivity bridge Model GC-16B2 (Beckman Instruments, Inc., Fullerton, California) following the method described in the operating manual.

Turbidity

The turbidity of the samples was determined using a HF Instruments Model DRT 100 Turbidimeter as per the instruction manual.

Total Organic Carbon (TOC)

The TOC was determined using a Beckman Model 915 TOC analyzer (Beckman Instruments, Inc., Fullerton, California) following the method described in the operating manual.

Filtered Organic Carbon (FOC)

1) Filter sample through a prewashed Whatman GFC glass fiber filter.

2) Determine TOC using a Beckman Model 915 TOC analyzer (Beckman Instruments, Inc. Fullerton, California) following method described in operating manual.

Total Suspended Oil (TSO) and Total Suspended Solids (TSS)

1) Filter known volume (V) of sample through a washed, ashed, and preweighed (W_1) Whatman glass fiber filter.
2) Dry filter overnight at 100°C.
3) Weigh filter (W_2).
4) Ash filter. Care must be taken to increase temperature slowly from 150°C to 450°C to volatilize oil without burning. Muffle for 1 hour at 450°C.
5) Weigh filter (W_3)
6) Calculations:

$$TSO = \frac{W_2 - W_3}{V} \qquad TSS = \frac{W_3 - W_1}{V}$$

note : Weight loss at 100°C assumed to be only water - along with water loss there may be a loss of the low-boiling components of the oil - resulting in a low TSO value.

Total Dissolved Solids (TDS) and Total Dissolved Oil (TDO)

1) Weigh crucible (W_1).
2) Transfer all filtrate from TSS + TSO determination of known volume (V) to crucible.
3) Dry crucible overnight at 60°C, dessicate, and weigh (W_2).
4) Ash residual in crucible. Muffle for 1 hour at 450°C.
5) Dessicate and weigh crucible (W_3)

6) Calculations:

$$TDS = \frac{W_2 - W_1}{V} \qquad TDO = \frac{W_3 - W_1}{V}$$

Anion Scan (IC)

Samples were analyzed using Ion Chromatography for Cl, PO_4, SO_4, NO_3, NO_2, and F. Results were reported in mg/L.

Cation Scan (DCP)

The samples were analyzed by direct coupled plasma for Ag, Al, Ba, Be, Ca, Cd, Co, Cr, Cu, Fe, K, Mg, Mn, Mo, Na, Ni, Pb, Sr, V, Zn. Results were reported in mg/L.

Silica

Silica is determined on a filtered (0.45 micron) aliquot following the procedure described in AWWA Standard Methods 15/B (Molybdosilicate Method). Accurate determination of total silica in the raw emulsion was impossible as a result of interference from oil and other components.

Calcium, Magnesium

1) Fitler sample through a prewashed Whatman GFC glass fiber filter.
2) Determine metals concentration using a Jarrel-Ash Atomic Absorption Spectrometer Model 82-500 (Jarrel-Ash Inc. Waltham, Mass.) following methods described in operating

manual (Calcium Analytical Method No. 9 Magnesium Analytical Method No. 18).

Calculated Hardness

Total hardness is calculated by multiplying the concentration of Ca and Mg by the proper factor to obtain equivalent calcium carbonate, and by summing these $CaCO_3$ concentrations. To obtain the $CaCO_3$ equivalent (mg/L) of Ca and Mg, the concentrations were multiplied by 2.5 and 4.116 respectively.

Oil, Grease, and Ethyl Acetate Extractable Surfactants

Attempts to extract oil and grease using freon were unsuccessful. A method was developed which involved breaking the emulsion and analyzing the separate components. A 200 mL aliquot of sample was frozen and allowed to thaw. The entire sample, except for a small amount of oil stuck to the inside of the container, was filtered through a No. 40 ashless filter. The solids portion on the filter was extracted with approximately 300 mL of freon as per the standard method for oil and grease determination, (AWWA Standard Methods, Procedure 503, Section C).

The filtrate was extracted 3 times, each time with 200 mL of ethyl acetate. During the third extraction, the pH was lowered to 1 with 1 mL of 1+1 H_2SO_4. The water phase from this extraction was separated and extracted with 50 mL of ethyl acetate. An additional 200 mL of ethyl acetate was added to the non aqueous phase to redissolve solids formed during the pH adjustment. The solvents were combined and residual water was removed by passing the sample through a filter of $NaSO_4$. The dried solvent was rotovaped and blown down with N_2 gas and weighed to determine the ethyl acetate extractable surfactant material in the oil remaining in the sample.

The oil remaining in the container was dissolved in toluene then blown down to determine the mass of this oil.

Surfactants for the petroleum industry

Organics Division

Witco Chemical Corporation

Reverse Demulsifiers

Witbreak RTC Oil-In-Water Demulsifiers

Oil-in-water emulsions or "reverse" emulsions occur naturally in oilfield production, especially in low-gravity crude producing areas. These types of emulsions may contain 0.5 to 1 percent recoverable oil. Reverse demulsifiers are used to resolve these emulsions in order to recover more crude and produce clean water for easy disposal.

Oil-in-water demulsifiers find use in several industrial processes such as:

Crude oil desalting —	Wash water must be treated to remove oil to permit easy disposal.
Industrial cleaning —	Solvent washing is used in certain industries and the solvent is subsequently removed in the process by a water/detergent solution. Solvents used may vary from hydrocarbons such as xylene or kerosene to chlorinated hydrocarbons. Resulting oil-in-water emulsions may present a disposal problem unless the oil is removed. Separation of the emulsions permits easy disposal of water and possible recycling of solvent.
Cutting oil emulsions —	Resolution of used cutting oil emulsions permits disposal of the water phase by conventional means.

Typical Properties

Witbreak RTC-323

Chemical Description	Polymeric Amine
Appearance	Liquid
pH, 5% in 3:1 Water/Isopropanol	9 - 11
Viscosity, Brookfield, at 25°C (77°F), cps	660 - 760
Pour Point, PMCC, °F	25
Flash Point, °F	>200
Density, lb/gal	8.5
Solubility, 10% in	
Water	Soluble
Xylene	Insoluble
Isopropanol	Insoluble
Kerosene	Insoluble

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APPENDIX B

RESULTS OF EMULSION CHARACTERIZATION

RESEARCH
FOUNDATION

SHERIDAN PARK RESEARCH COMMUNITY
MISSISSAUGA, ONTARIO, CANADA [illegible] • (416) 822-41[illegible] • TELEX 06-982311

MATERIALS DIVISION

Non-Metallic Minerals Centre

PARTICLE SIZE ANALYSIS OF A OIL/CLAY SUSPENSION

Fine Particle Laboratory Services
Report No. FPL 50005-84-19

M.P. Haley
May 28,1984

for

Zenon Environmental
845 Harrington Court
Burlington, Ontario
L7N 3P3

Attention: Ms. Chris Zienkowicz

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1. INTRODUCTION

One sample of a oil/clay suspension was submitted for particle size analysis of both the oil and clay fractions.

2. METHOD

The initial methodology to separate and analyze the two distinct fractions was to involve a Coulter Counter particle sizing on the raw material, followed by filtering a metered sample which could be collected, washed and analyzed using optical microscopy/image analysis. Unfortunately, the portion collected on the filter paper contained a large number of dark indefinite shapes which were attributed to the oil droplets present in the sample. By washing a portion of the sample with toluene in a separatory funnel, it was noted that the oil fraction could be separated thereby leaving the clay particles in the aqueous phase which could be collected on a filter. Following this procedure, however, posed certain problems in developing sufficient contrast of the clay particles to enable proper detection with the Quantimet image analysing computer. An alternate procedure was then attempted.

Since it had been shown that toluene was effective in removing the oil fraction from the clay particles, a 1 mL portion of the raw sample was washed with 50 mL of toluene in a separatory funnel. To this, 9 mL of Isoton (a particle free aqueous electrolyte) was added to aid in particle dispersion and separation. Once the oil had been removed from the aqueous phase (as noted by change in color of both the toluene and aqueous layers), the aqueous portion was collected for particle size analysis. Residual clay particles in the separatory funnel were washed through and collected with successive additions of Isoton. This sample was then analyzed on the Coulter Counter TAII.

Particle size measurement on both the raw sample and the toluene washed sample was conducted on the Coulter Counter TAII using a 70 µm aperture tube which provided an effective size range of 1.6 - 32+ µm. By preparing accurately diluted samples and metering a precise sample volume through the aperture tube, estimates of particle concentration, within the range stated above, could be deduced. To arrive at the size distribution for the oil droplets, the toluene washed sample data (e.g. clay particles) were subtracted from the total or raw sample.

3. RESULTS

Concentration estimates are presented in Table I while the size distribution (by volume and number) for the total, clay, and oil fractions are presented in the Appendices I - III.

TABLE I

Fraction	Particle/mL Solution
Raw	474.4×10^4
Clay (Toluene Washed)	34.1×10^4
Oil (Raw - Clay)	440.3×10^4

4. DISCUSSION

From the results of Table I, it appears that the majority of the "particles" observed in the raw solutions, are in fact oil droplets (~93%). The mean particle size for these oil droplets appears to be approximately 7.8 µm while that of the clay particulates is 9.8 µm.

Mike P. Haley
Mike P. Haley
Assistant Research Scientist

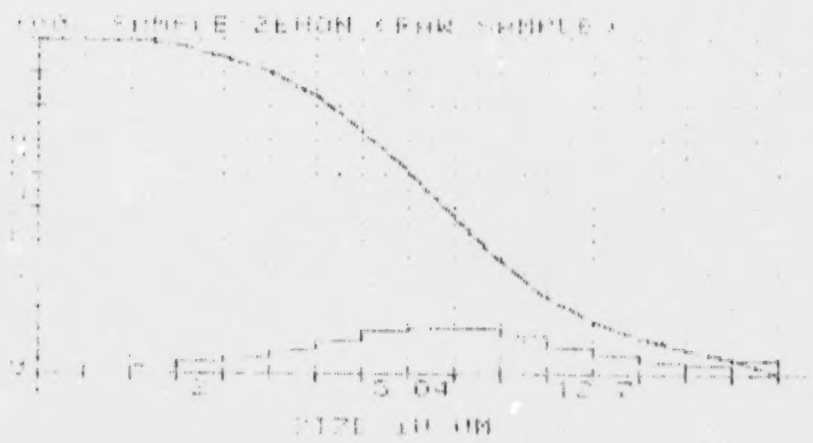
Ivan H. Joyce, Manager
Non-Metallic Minerals Centre

/gb

```
DATE: 25 MAY 1984
SAMPLE: ZENON (RAW SAMPLE)
ELECTROLYTE: ISOTON
DISPERSANT: NIL
EQUIPMENT: TA II
APERTURES: 70
OPERATOR: MFH
COMMENTS: DILUTED 4[illegible]
*****************************************
CH. #   SIZE          DIFF        CUM
                      VOL %       VOL %
*****************************************
1       1             0           100
2       1.26          0           100
3       1.59          1.8         100
4       2             1.4         98.2
5       2.52          4.8         96.81
6       3.17          [illegible] [illegible]
7       4             8.19        [illegible].82
8       5.04          12.35       [illegible]
9       6.35          13.45       [illegible].84
10      8             13.35       4[illegible]
11      10.08         11.[illegible] [illegible]
12      12.7          7.45        [illegible].68
13      16            5.09        15.12
14      20.16         3.6         10.0[illegible]
15      25.4          2.8         6.45
16      32            3.7         3.[illegible]


SAMPLE ZENON (RAW SAMPLE)10:[illegible] RAW
*****************************************
          VOLUME % STATISTICS
*****************************************

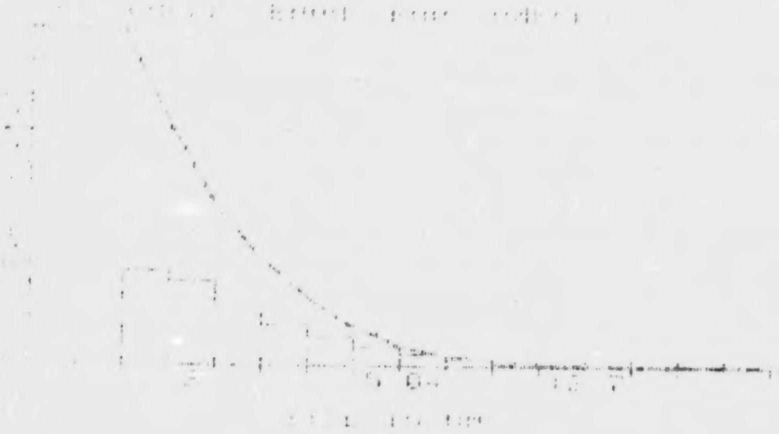
MEAN:                     7.79 UM
MEDIAN:                   7.62 UM
MODE                      7.85 UM
STANDARD DEVIATION:       2 UM
COEF. VAR.:               25.73%
SKEWNESS:                 1.06 NEGATIVE
KURTOSIS:                 1.86 PLATYKURTIC
```



```
DATE: 2[illegible] MAY 1984
SAMPLE: ZENON (RAW SAMPLE).
ELECTROLYTE: ISOTON
DISPERSANT: NIL
EQUIPMENT. TA II
APERTURES: 70
OPERATOR: MFH
COMMENTS: DILUTED 400X
****************************************
CH.#    SIZE             DIFF        CUM
                         POP %       POP %
****************************************
1       1                0           100
2       1.26             0           100
3       1.59             2[illegible].61     100
4       2                24.46       [illegible]
5       2.52             15.78       47.91
6       3.17             11.87       31.53
7       4                8.57        1[illegible].66
8       5.04             5.41        11.07
9       6.35             3.08        5.[illegible]4
10      8                1.56        [illegible]
11      10.08            .85         1
12      12.7             [illegible]         [illegible]
13      16               .07         .11
14      20.16            [illegible]         .04
15      25.4             .01         .02
16      [illegible]              .0[illegible]         .0[illegible]

SAMPLE:ZENON (RAW SAMPLE) ID:ZENON RAW
****************************************
         POPULATION STATISTICS
****************************************

MEAN                     2.78 UM
MEDIAN                   2.47 UM
MODE                     1.93 UM
STANDARD DEVIATION       1.55 UM
COEF. VAR.               55.83%
SKEWNESS                 1.1 POSITIVE
KURTOSIS                 1.16 PLATYKURTIC
```



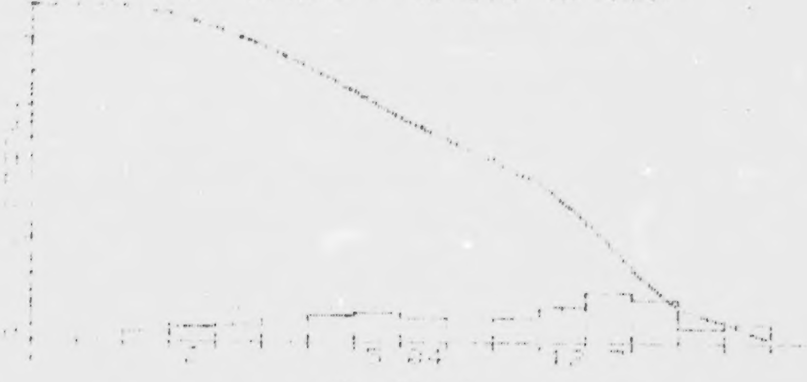
ORF

```
DATE: 25 DEC 1984
SAMPLE: ZENON (TOLUENE WASHED)
ELECTROLYTE: ISOTON
DISPERSANT: OIL
EQUIPMENT: TA II
APERTURES: 70
OPERATOR: MRH
COMMENTS: DILUTED 1:??
*****************************************
CH.#   SIZE         DIFF       CUM
                    VOL %      VOL %
*****************************************
1      1            0          100
2      1.26         0          100
3      1.59         ..         100
4      2            4.1        97.3
5      2.52         5          93.21
6      3.17         6.29       88.21
7      4            7.34       81.92
8      5.04         6.04       74.57
9      6.35         6.59       68.43
10     8            5.89       59.84
11     10.08        6.75       53.73
12     12.7         10.59      47.15
13     16           14.59      36.56
14     20.16        12.57      21.98
15     25.4         5.8        9.37
16     32           5.54       3.54


SAMPLE1:ZENON (TOLUENE WASHED) ID:ZENON WASHED
*****************************************
        VOLUME % STATISTICS
*****************************************

MEAN:                     9.76 UM
MEDIAN:                   11.53 UM
MODE:                     10.66 UM
STANDARD DEVIATION:       2.28 UM
COEF. VAR.:               23.39%
SKEWNESS:                 .83 NEGATIVE
KURTOSIS:                 2.54 PLATYKURTIC
```

VOL. SAMPLE: ZENON (TOLUENE WASHED)

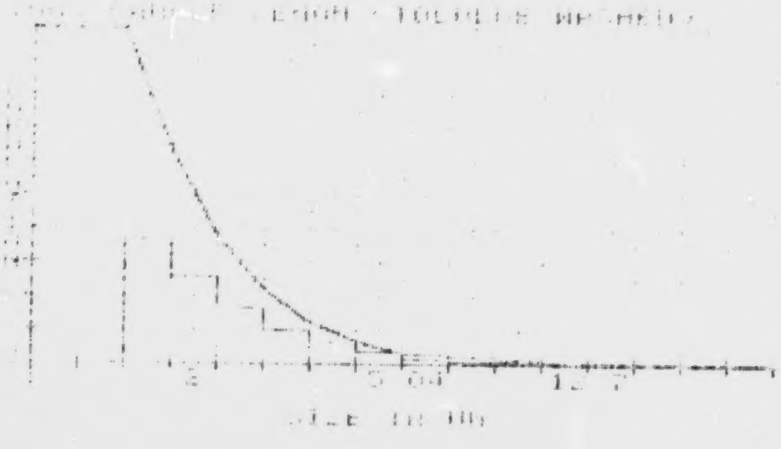


SIZE IN UM

```
DATE: 25 MAY 1984
SAMPLE: ZENON (TOLUENE WASHED)
ELECTROLYTE: ISOTON
DISPERSANT: NIL
EQUIPMENT: TA II
APERTURES: [illegible]
OPERATOR: MPH
COMMENTS: DILUTED [illegible]
*****************************************
CH.#    SIZE          DIFF        CUM
                      POP %       POP %
*****************************************
1       1             0           100
2       1.26          0           100
3       1.59          36.67       100
4       2             25.8        63.33
5       2.52          15.63       37.53
6       3.17          [illegible] 21.9
7       4             5.61        11.8
8       5.04          3.15        6.19
9       6.35          1.78        3.04
10      8             .57         1.26
11      10.08         .34         .69
12      12.7          [illegible] .35
13      16            .16         .28
14      20.16         .06         .12
15      25.4          .01         .06
16      32            .01         .01


SAMPLE: ZENON (TOLUENE WASHED)1: ZENON WASHED
*****************************************
        POPULATION STATISTICS
*****************************************

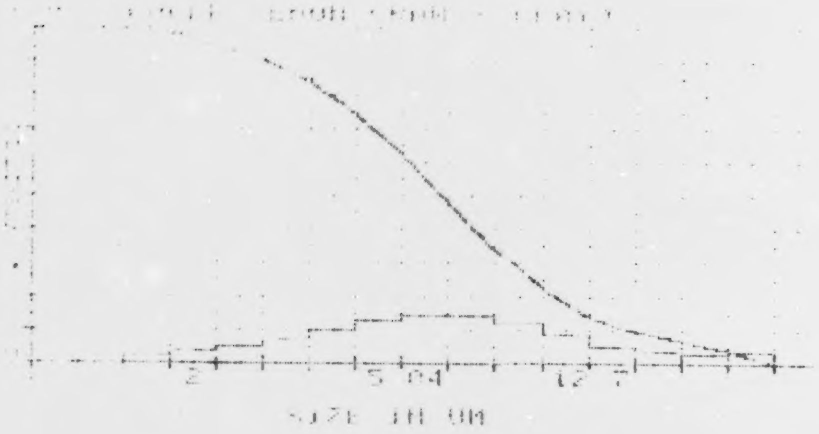
MEAN:                   2.5 UM
MEDIAN:                 2.25 UM
MODE                    1.9 UM
STANDARD DEVIATION:     1.48 UM
COEF. VAR.:             59.18%
SKEWNESS:               1.1 POSITIVE
KURTOSIS:               1.17 PLATYKURTIC
```



```
DATE. 25 MAY 1984
SAMPLE: ZENON (RAW - CLAY)
ELECTROLYTE: ISOTON
DISPERSANT: NIL
EQUIPMENT: TA II
APERTURES: 70
OPERATOR: MFH
COMMENTS: CALCULATED DIFFERENCE
****************************************
CH.#   SIZE          DIFF       CUM
                     VOL %      VOL %
****************************************
1      1             0          100
2      1.26          0          100
3      1.59          1.8        100
4      2             3.2        98.2
5      2.52          4.5        94.9
6      3.17          6.5        90.4
7      4             9.6        83.9
8      5.04          12.3       74.3
9      6.35          14.1       62
10     8             14.7       47.9
11     10.08         11.5       33.2
12     12.7          5.1        [illegible]
13     16            4.4        16.6
14     20.16         3          [illegible]
15     25.4          1.[illegible]  [illegible]
16     32            [illegible]    [illegible]

SAMPLE:ZENON (RAW - CLAY)ID.ZENON DIFF
****************************************
       VOLUME % STATISTICS
****************************************

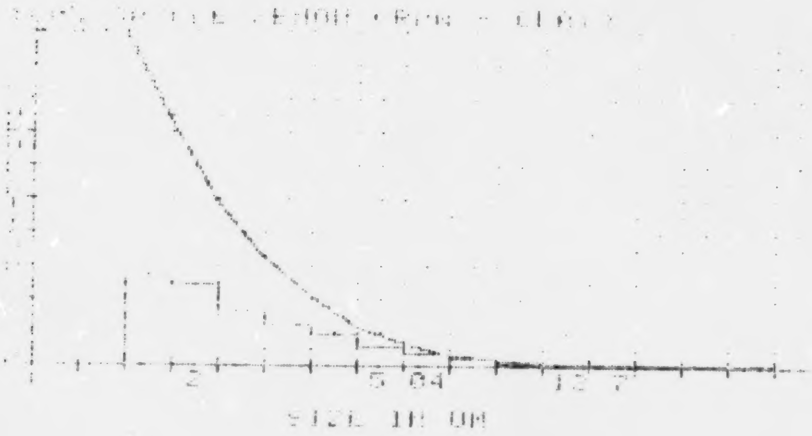
MEAN:                    7.81 UM
MEDIAN:                  7.72 UM
MODE                     8.14 UM
STANDARD DEVIATION:      1.97 UM
COEF. VAR.:              25.24%
SKEWNESS:                1.05 NEGATIVE
KURTOSIS:                1.8 PLATYKURTIC
```



```
DATE: 25 MAY 1984
SAMPLE: ZENON (RAW - CLAY)
ELECTROLYTE: ISOTON
DISPERSANT: NIL
EQUIPMENT: TA II
APERTURTES: 70
OPERATOR: MPH
COMMENTS: CALCULATED DIFFERENCE
****************************************
CH.#    SIZE          DIFF        CUM
                      POP %       POP %
****************************************
1       1             0           100
2       1.26          0           100
3       1.59          26.91       100
4       2             24.38       73.09
5       2.52          16.42       48.71
6       3.17          12.02       32.29
7       4             8.81        20.27
8       5.04          5.6         11.46
9       6.35          3.21        5.86
10      8             1.63        2.64
11      10.08         .68         1.01
12      12.7          .23         .33
13      16            .06         .1
14      20.16         .02         .04
15      25.4          .01         .02
16      32            .01         .01


SAMPLE:ZENON (RAW - CLAY)ID:ZENON DIFF
****************************************
        POPULATION STATISTICS
****************************************

MEAN:                       2.6 UM
MEDIAN:                     2.49 UM
MODE                        1.76 UM
STANDARD DEVIATION:         1.35 UM
COEF. VAR.:                 55.51%
SKEWNESS:                   1.1 POSITIVE
KURTOSIS:                   1.15 PLATYKURTIC
```



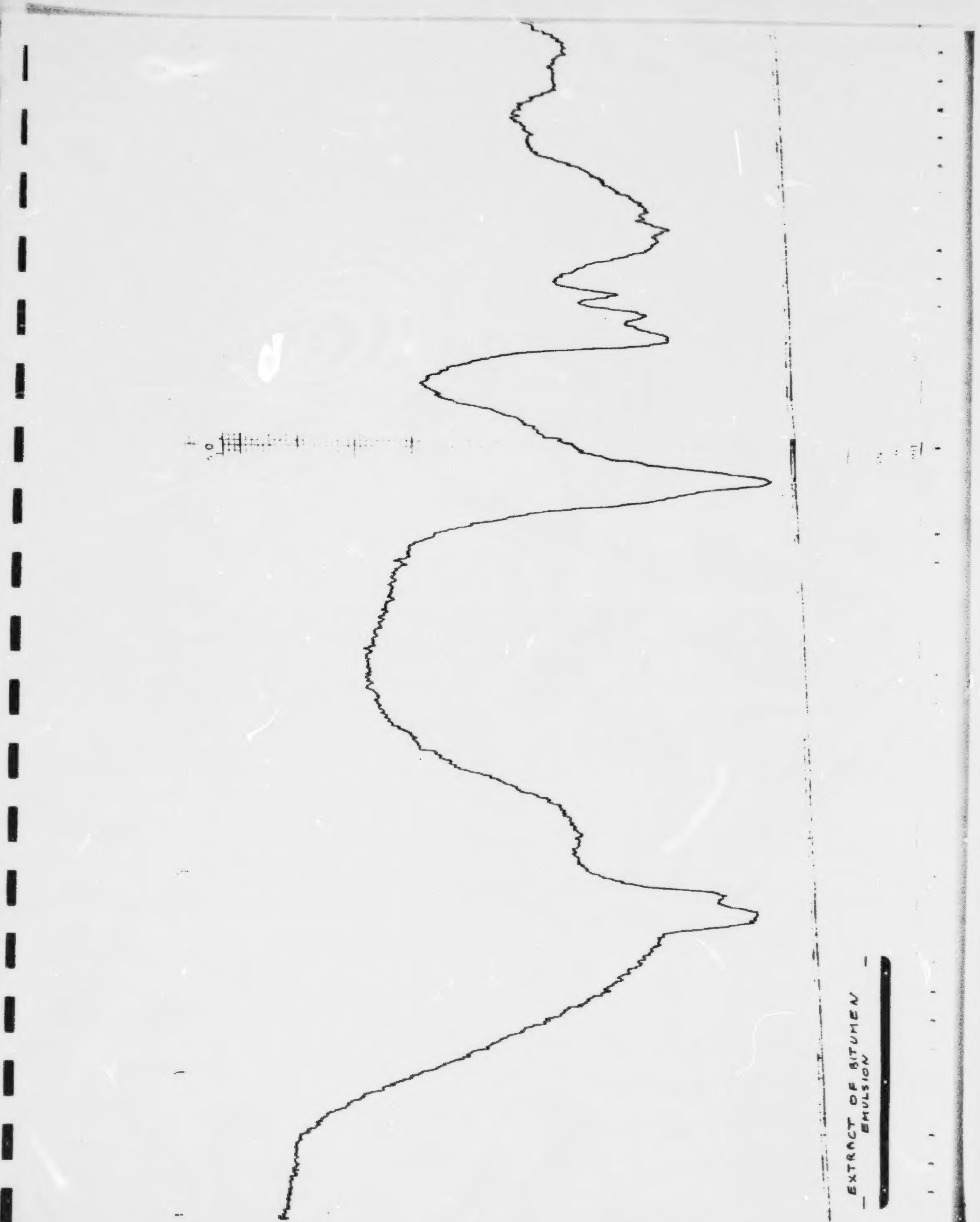
ORF

INFRA RED SPECTROPHOTOMETRY

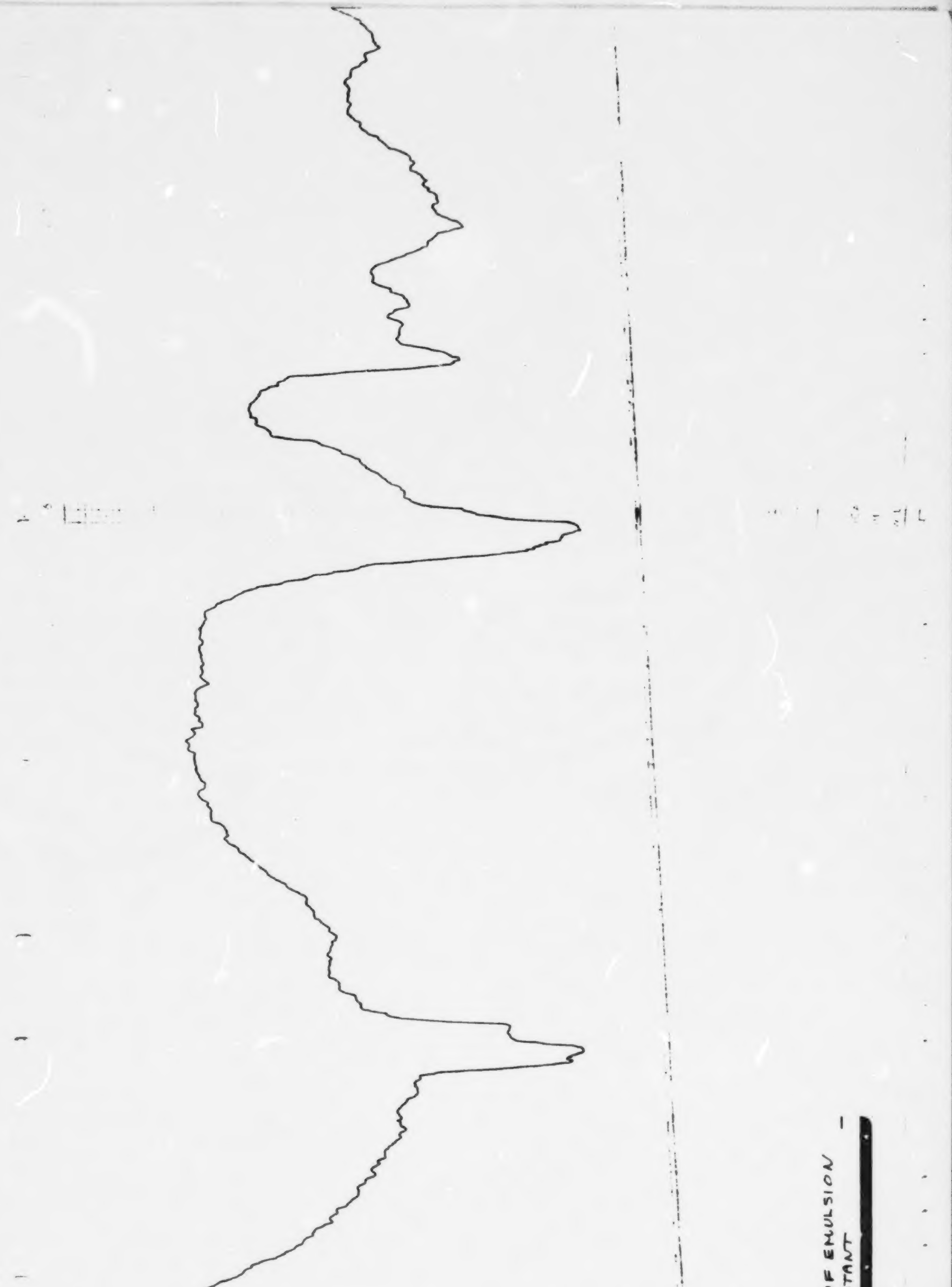
Conditions

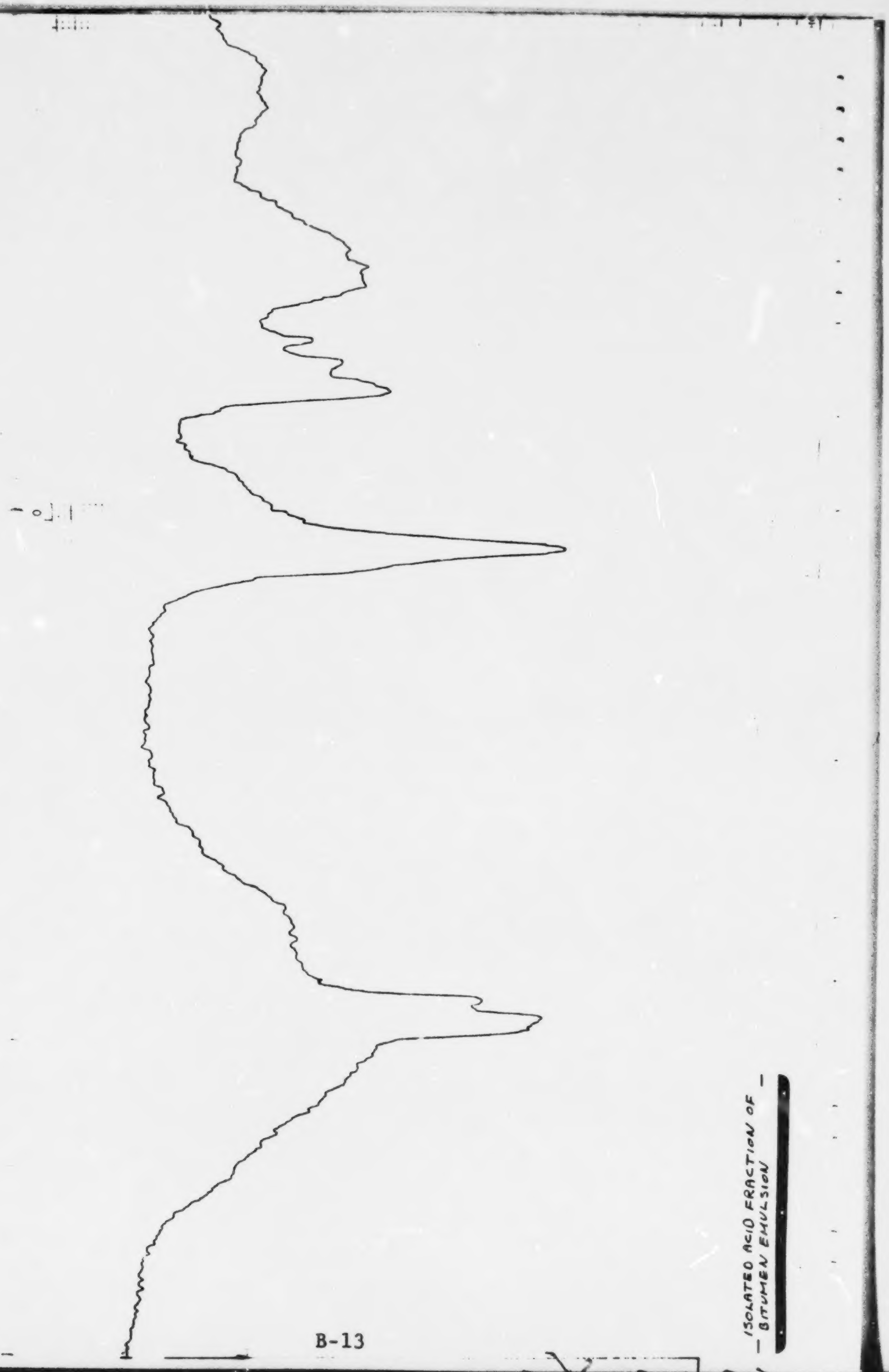
Instrument -	Perkin Elmer 283
Analysis -	KBr Plates
Mode -	% Transfer
Scan -	4000 - 200 CM^{-1}
Scan Rate -	3 minutes
Calibration -	Polystyrene

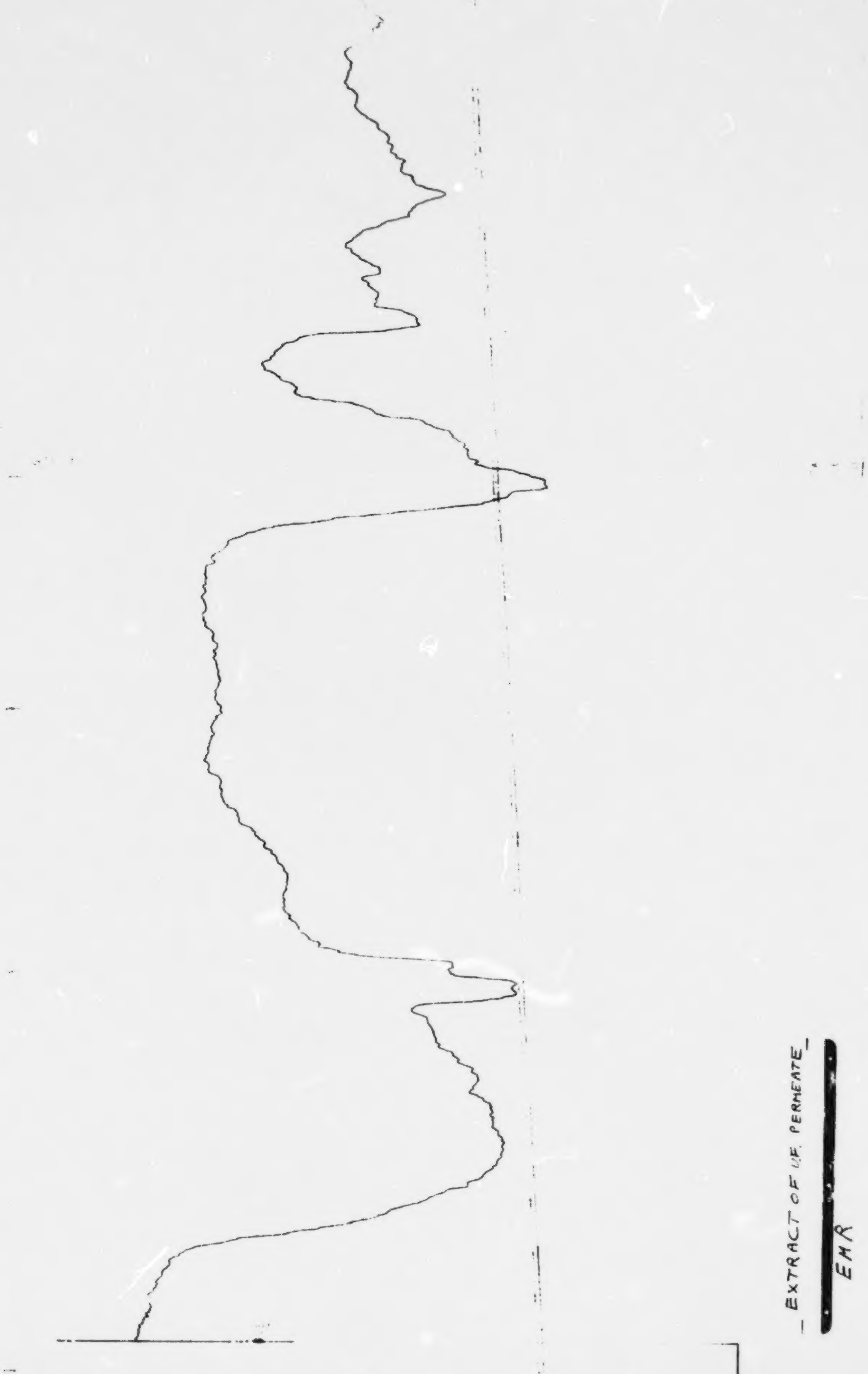
Results as follow

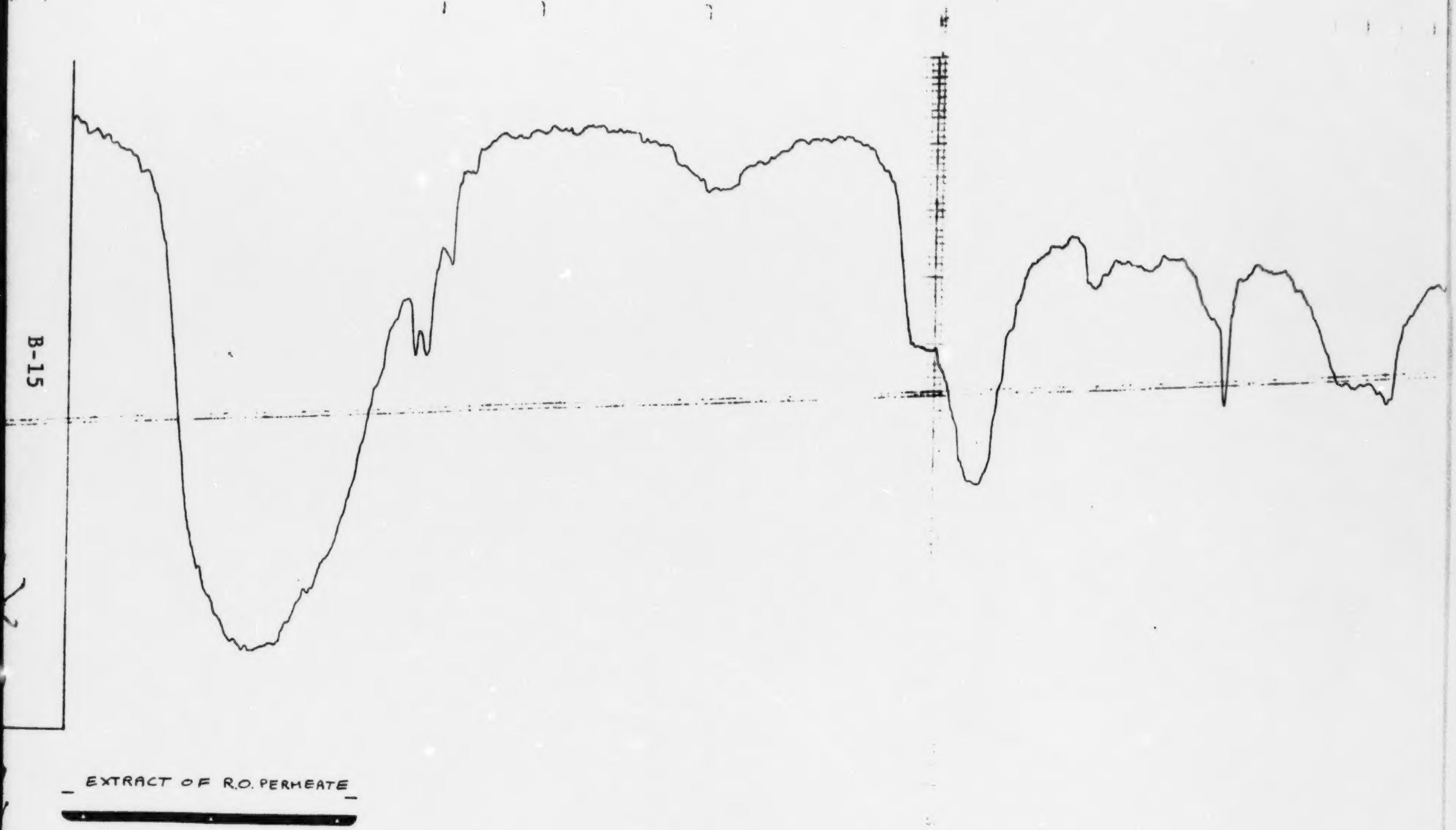


EXTRACT OF BITUMEN
EMULSION


EXTRACT OF EMULSION SUPERNATANT
EMR


ISOLATED ACID FRACTION OF
BITUMEN EMULSION


EXTRACT OF U.F. PERMEATE
EMR


EXTRACT OF R.O. PERMEATE

Gas Chromatograph Analysis

Conditions

Instrument - HP5880-Level IV

Injection - Splitless, 30 sec.

Injector Temperature - 250°c

Split - 40 mL/min

Septum Flush - 6 mL/min

Column - 60 M X 0.31 mm ID DB-5

Oven Temperature Profile - 60°c for 2 min, up to 220°c at 8°/min, then up to 270°c at 12°/min

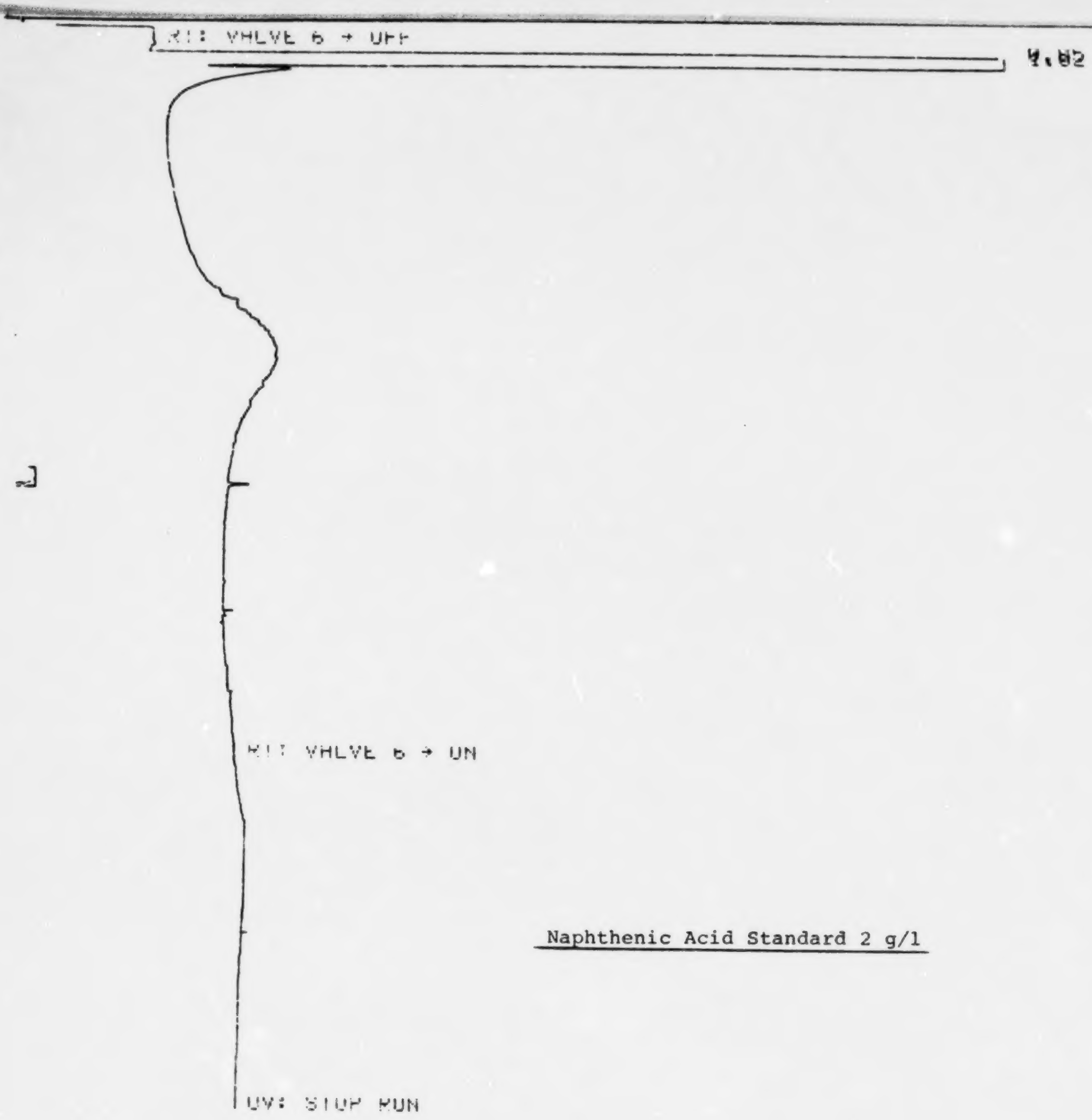
Detection - Flame Ionization

Detector Air Flow - 450 mL/min

Detector Hydrogen Flow - 30 mL/min

Detector Makeup - Nitrogen @ 30 mL/min

Results as Follows



Naphthenic Acid Standard 2 g/l

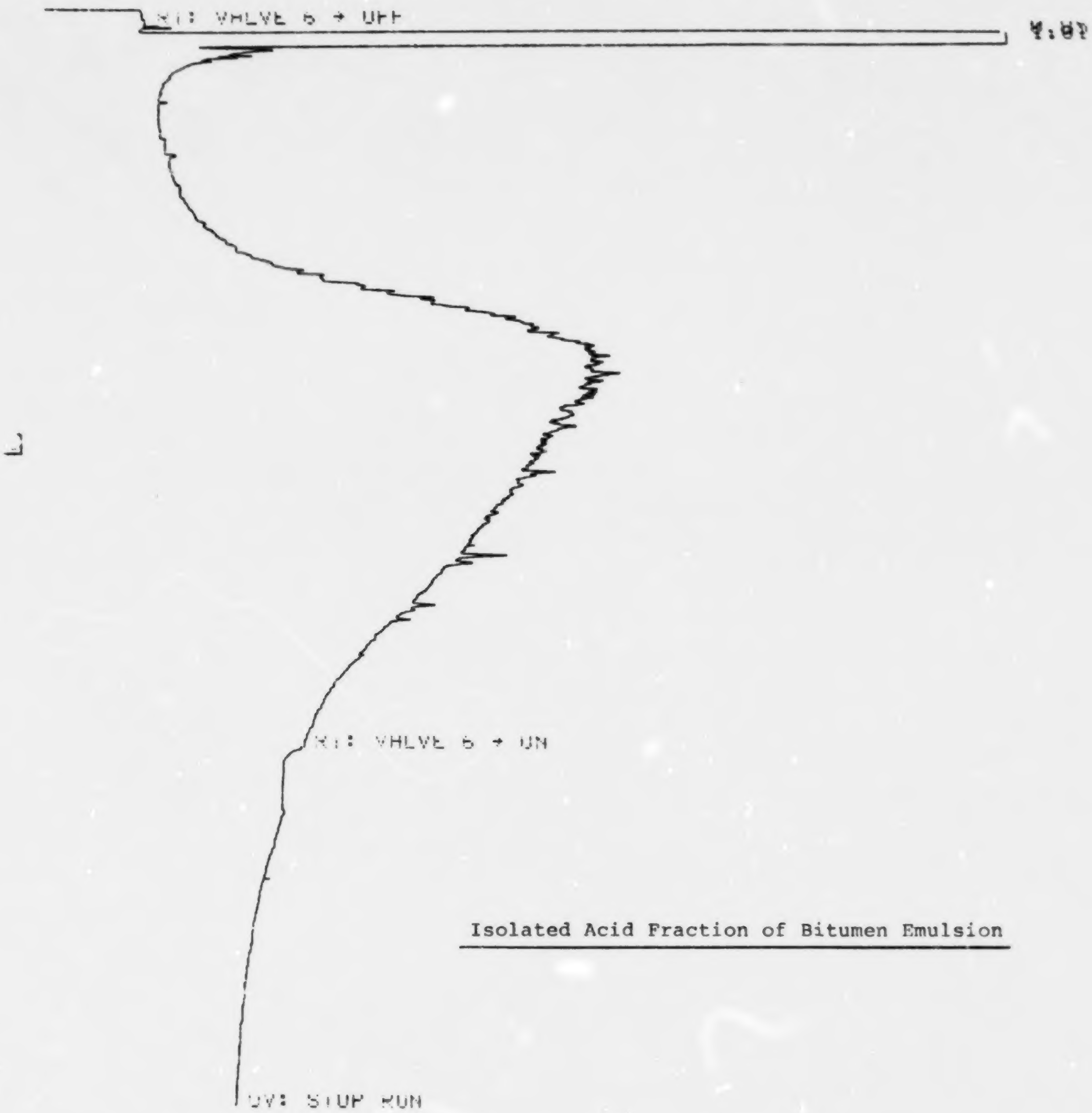
```
HP 5880A  SAMPLER INJECTION @ 17:21  NOV 28, 1984
  SAMPLE # :  ID CODE   :
        10   2MG/ML TSA  in MeCl2
AREA %

   RT                  AREA  TYPE    AREA %

  0.85        491452.00   BH      99.512
  1.02          2409.54   HH       0.488

TOTAL AREA = 493861.00
MULTIPLIER = 1

VALVE 6 OFF
```



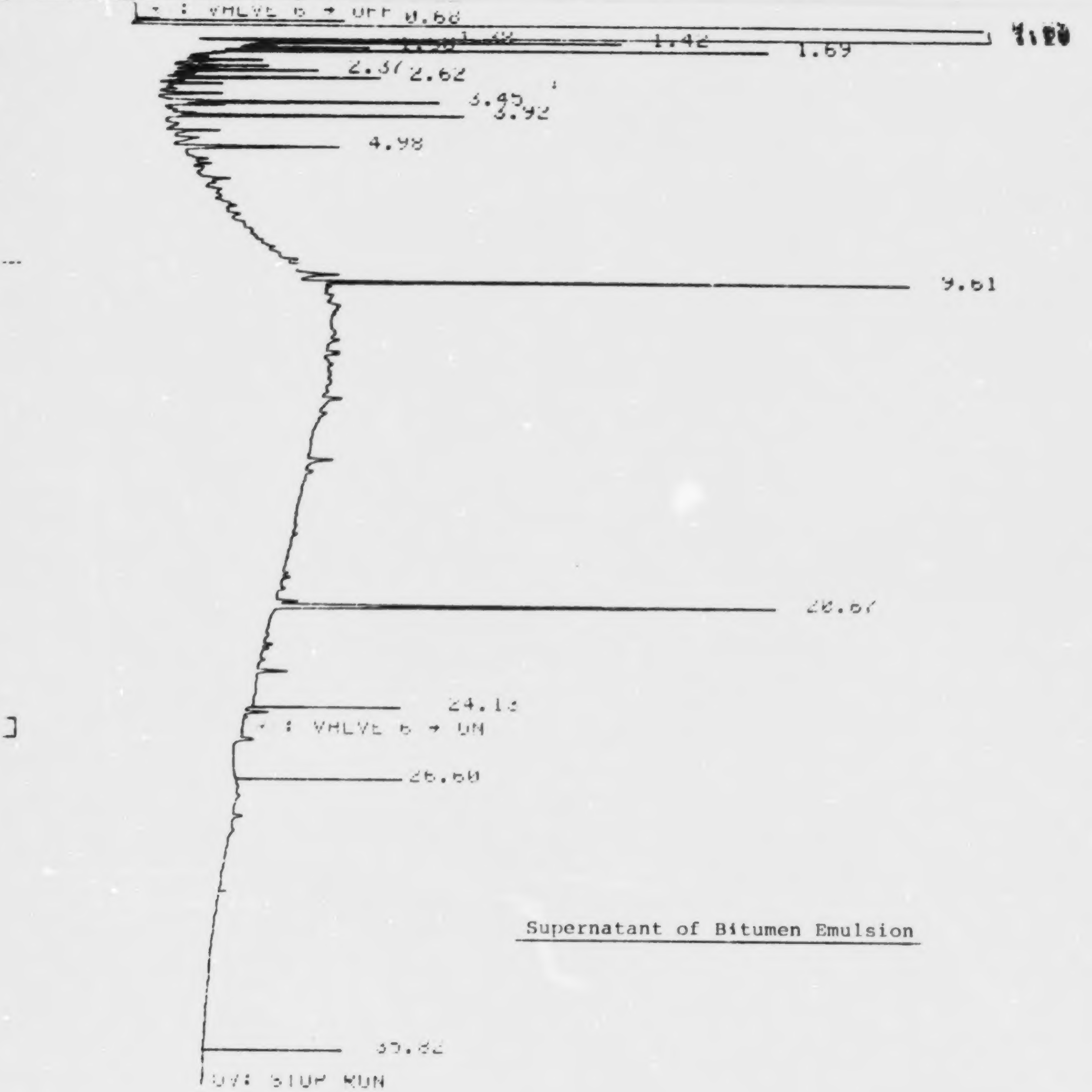
Isolated Acid Fraction of Bitumen Emulsion

```
[HP] 5880A  SAMPLER INJECTION @ 15:09  NOV 28, 1984
  SAMPLE # :  ID CODE   :
          9    BIT EMUL2
AREA %

   RT               AREA  TYPE    AREA %

  0.65       545132.00   BH     99.401
  1.01         3283.79   HH      0.599

TOTAL AREA = 548415.00
MULTIPLIER = 1
```

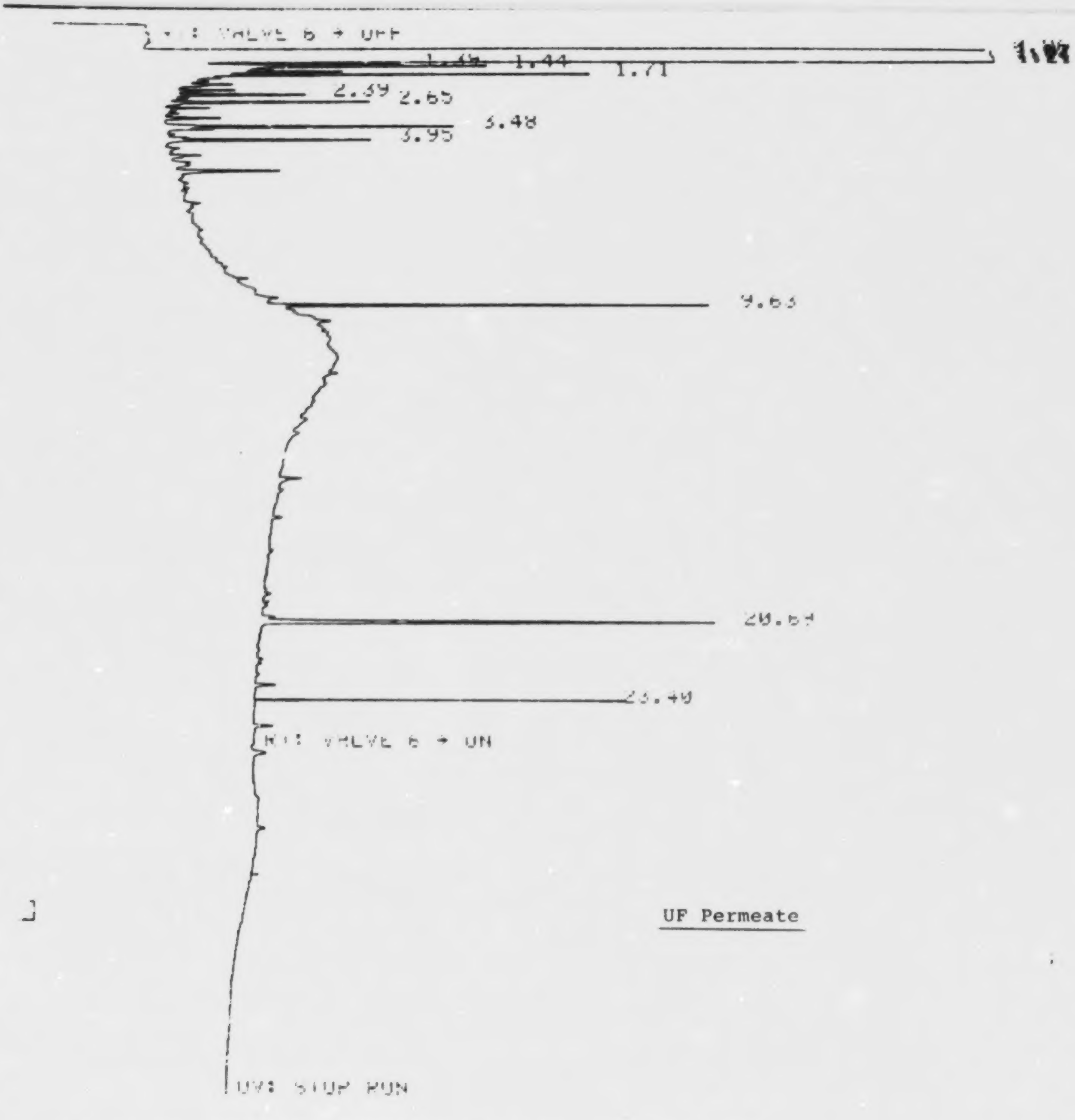


Supernatant of Bitumen Emulsion

```
&HP# 5880A  SAMPLER INJECTION @ 14:26  NOV 28, 1984
   SAMPLE # :   ID CODE    :
          /     BIT SUPER
AREA %

   RT              AREA   TYPE    AREA %

   0.68             7.33   BP      0.001
   0.85        487222.00   BV     99.375
   1.01          2070.84   VV      0.422
   1.06           135.01   VV      0.028
   1.10            20.22   BP      0.004
   1.20           189.45   VP      0.039
   1.33            27.04   PV      0.006
   1.42            52.74   VV      0.011
   1.58            21.75   VV      0.004
   1.69            71.46   VV      0.015
   2.37            24.58   VV      0.005
   2.62            34.-    PV      0.007
   3.45            5        V      0.012
```

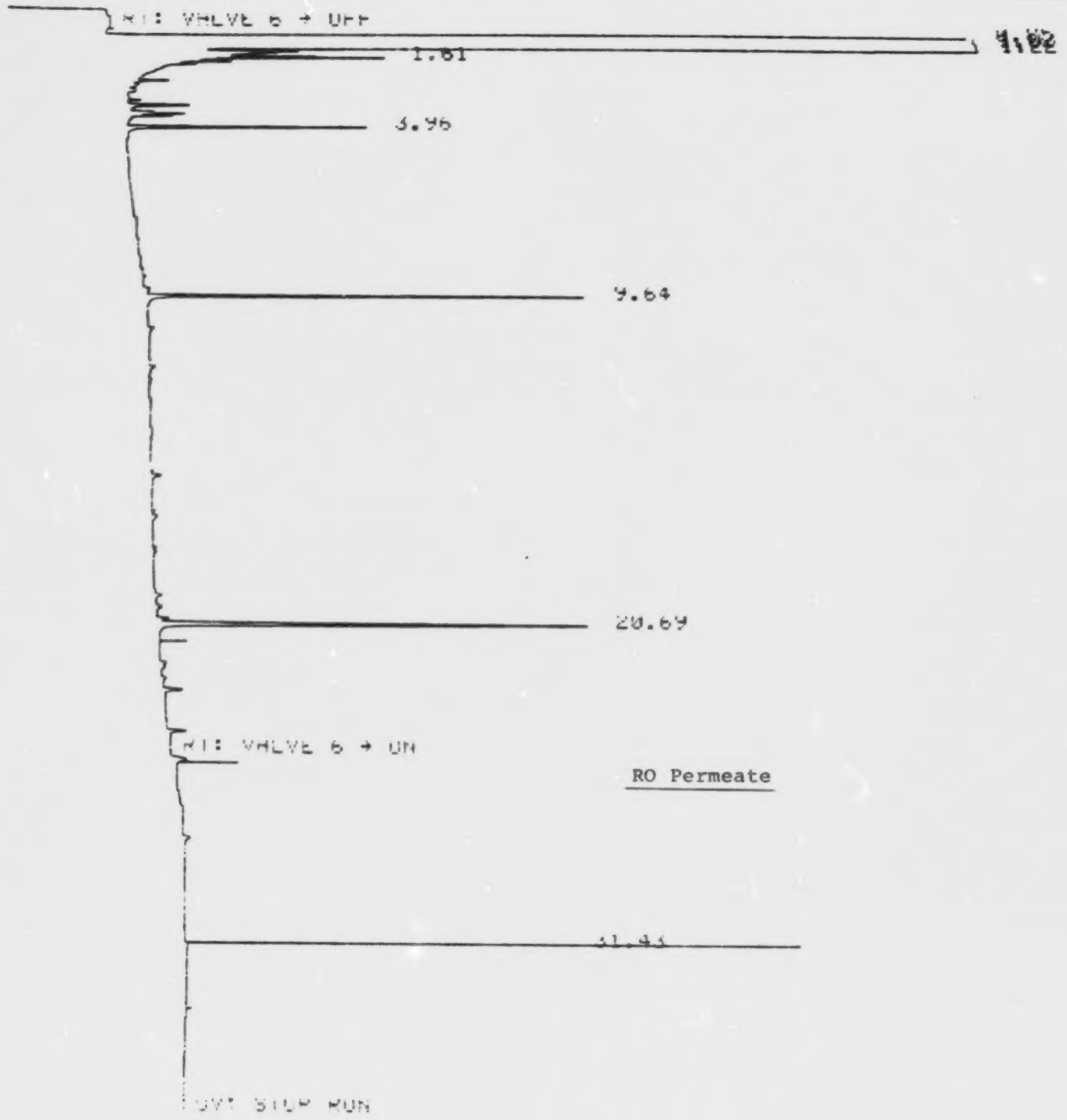


UF Permeate

```
RUN # [illegible]   SAMPLER INJECTION @ 13:42  NOV 28, 1984
  SAMPLE # :  ID CODE   :
          5       UF PERM
AREA %

   RT                 AREA   TYPE    AREA %

   0.86          467628.00    BV     98.957
   1.02            2858.24    VV      0.605
   1.07             475.23    VV      0.101
   1.11             502.89    VV      0.106
   1.21             541.38    VV      0.115
   1.39              41.81    VV      0.009
   1.44              60.84    VV      0.013
   1.71              71.86    VV      0.015
   2.39              26.56    VV      0.006
   2.65              31.97    BV      0.007
   3.48              62.09    BV      0.013
   3.95              36.53    VB      0.008
```



RO Permeate

```
RUN # [illegible]   SAMPLER INJECTION @ 12:58  NOV 28, 1984
  SAMPLE # :  ID CODE   :
         5        RO PERM
AREA %

  RT              AREA  TYPE    AREA %

  0.85       558864.00   BV     98.009
  1.02         2904.56   VV      0.509
  1.07          969.25   VV      0.173
  1.22         7169.52   VV      1.257
  1.51           36.07   VV      0.006
  3.96           39.76   BB      0.007
  9.64           75.59   BB      0.013
 20.69          124.17   BB      0.022
 31.43           11.41   BB      0.002

TOTAL AREA = 570214.00
MULTIPLIER = 1
```

APPENDIX C

RESULTS OF PRESCREENING TESTS

PURE WATER FLUX RATES
PRESSURE 100 psi TEMP 20 C

CELLULOSE #12

TIME	MLS	MLS/MIN	M3/M2/D
15.0	9.0	0.60	0.76
31.0	5.6	0.35	0.44
46.0	4.0	0.27	0.34
61.5	5.1	0.33	0.41
76.5	4.9	0.33	0.41

CELLULOSE ACETATE OK

TIME	MLS	MLS/MIN	M3/M2/D
15.0	40.5	2.70	3.40
30.0	40.0	2.67	3.36
45.0	40.0	2.67	3.36
60.0	39.5	2.63	3.32
75.0	40.0	2.67	3.36

CELLULOSE #13

TIME	MLS	MLS/MIN	M3/M2/D
15.0	64.0	4.27	5.38
30.0	56.0	3.73	4.70
45.0	18.0	1.20	1.51
60.0	16.0	1.07	1.34

CELLULOSE ACETATE 20K

TIME	MLS	MLS/MIN	M3/M2/D
15.0	71.5	4.77	6.01
30.0	72.0	4.80	6.05
45.0	72.0	4.80	6.05
60.0	72.0	4.80	6.05
75.0	72.0	4.80	6.05

CELLULOSE #23

TIME	MLS	MLS/MIN	M3/M2/D
15.0	19.0	1.27	1.60
31.0	11.0	0.69	0.87
46.0	21.0	1.40	1.76
61.5	27.0	1.74	2.19
76.5	26.0	1.73	2.18

CELLULOSE ACETATE 50K

TIME	MLS	MLS/MIN	M3/M2/D
15.0	19.5	1.30	1.64
30.0	21.5	1.43	1.81
45.5	22.0	1.42	1.79
61.0	21.0	1.35	1.71
76.0	20.0	1.33	1.68

CELLULOSE RC-51

TIME	MLS	MLS/MIN	M3/M2/D
15.0	1.7	0.11	0.14
30.0	1.9	0.13	0.16
45.0	2.2	0.15	0.18
60.0	2.1	0.14	0.18
90.0	3.9	0.13	0.16
120.0	4.0	0.13	0.17

CELLULOSE ACETATE CA-UF

TIME	MLS	MLS/MIN	M3/M2/D
15.0	17.0	1.13	1.43
30.0	16.8	1.12	1.41
45.0	16.8	1.12	1.41
60.0	16.8	1.12	1.41
75.0	16.4	1.09	1.38

PURE WATER FLUX RATES

PRESSURE 100 psi TEMP 20 C

T.F.C. G5

TIME	MLS	MLS/MIN	M3/M2/D
15.0	3.9	0.26	0.33
30.0	3.9	0.26	0.33
45.0	NA	0.00	0.00
60.0	3.4	0.23	0.29
75.0	3.5	0.23	0.29

T.F.C. G10

TIME	MLS	MLS/MIN	M3/M2/D
30.0	28.5	0.95	1.20
60.0	29.5	0.98	1.24
75.0	15.5	1.03	1.30
90.0	15.0	1.00	1.26
105.0	14.5	0.97	1.22

T.F.C. G20

TIME	MLS	MLS/MIN	M3/M2/D
30.0	71.0	2.37	2.98
60.0	74.0	2.47	3.11
75.0	37.5	2.50	3.15
90.0	37.5	2.50	3.15
105.0	37.0	2.47	3.11

T.F.C. G50

TIME	MLS	MLS/MIN	M3/M2/D
10.0	94.5	9.45	11.91
25.0	133.5	8.90	11.21
40.0	132.4	8.83	11.12
55.0	131.5	8.77	11.05
70.0	131.0	8.73	11.00

POLYSULFONE 0K

TIME	MLS	MLS/MIN	M3/M2/D
15.0	169.5	11.30	14.24
30.0	152.0	10.13	12.77
45.5	151.0	9.74	12.27
61.0	140.0	9.03	11.38
76.0	128.5	8.57	10.79

POLYSULFONE 20K

TIME	MLS	MLS/MIN	M3/M2/D
10.0	142.5	14.25	17.96
20.0	138.0	13.80	17.39
30.0	130.0	13.00	16.38
40.0	126.0	12.60	15.88
50.0	123.0	12.30	15.50

POLYSULFONE 50K

TIME	MLS	MLS/MIN	M3/M2/D
10.0	132.0	13.20	16.63
20.0	123.5	12.35	15.56
30.0	119.0	11.90	14.99
40.0	114.0	11.40	14.36
50.0	112.0	11.20	14.11

POLYSULFONE E500

TIME	MLS	MLS/MIN	M3/M2/D
15.0	231.0	15.40	19.40
30.0	214.0	14.27	17.98
45.5	190.5	12.29	15.49
60.0	173.5	11.97	15.08
75.0	171.0	11.40	14.36

PURE WATER FLUX RATES

PRESSURE 100 psi TEMP 20 C

C.H.P. Cat

TIME	MLS	MLS/MIN	M3/M2/D
10.0	205.0	20.50	25.83
25.0	270.0	18.00	22.68
40.0	228.0	15.20	19.15
55.0	175.0	11.67	14.70
70.0	180.0	12.00	15.12

C.H.P. B2

TIME	MLS	MLS/MIN	M3/M2/D
10.0	98.0	9.80	12.35
25.0	150.0	10.00	12.60
40.0	142.0	9.47	11.93
55.0	115.0	7.67	9.66
70.0	119.0	7.93	10.00

C.H.P. B3

TIME	MLS	MLS/MIN	M3/M2/D
15.0	183.0	12.20	15.37
30.0	154.0	10.27	12.94
45.0	154.0	10.27	12.94
60.0	147.0	9.80	12.35
75.0	143.0	9.53	12.01

POLYVINYLDIFLUORIDE

TIME	MLS	MLS/MIN	M3/M2/D
15.0	82.0	5.47	6.89
30.0	76.0	5.07	6.38
45.0	70.0	4.67	5.88
60.0	64.0	4.27	5.38
90.0	119.0	3.97	5.00
120.0	114.0	3.80	4.79

MEMBRANE SCREENING DATA
SALT WATER FLUX DATA   OPERATING PRESSURE 100 psi
BRACKISH WATER   TEMPERATURE 20 C

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
PS-0K	10.0	116.0	11.60	14.62	2049	2200	6.86
	21.5	130.0	11.30	14.24	2050	2200	6.82
	31.5	111.0	11.10	13.99	2030	2200	7.73

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
PS-20K	10.0	118.0	11.80	14.87	1990	2200	9.55
	21.5	131.0	11.39	14.35	1980	2200	10.00
	31.5	112.0	11.20	14.11	1990	2200	9.55

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
PS-50K	15.5	212.0	13.68	17.23	1960	2200	10.91
	30.5	194.0	12.93	16.30	1910	2200	13.18
	45.5	188.0	12.53	15.79	1950	2200	11.36

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
CELLULOSE	30.0	3.9	0.13	0.16	1170	2500	53.20
RC-51	60.0	3.9	0.13	0.16	2200	2475	11.11
	90.0	4.0	0.13	0.17	2325	2500	7.00
	120.0	4.0	0.13	0.17	2225	2500	11.00

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
PVDF	30.0	114.0	3.80	4.79	2150	2500	14.00
	60.0	112.0	3.73	4.70	2275	2475	8.08
	90.0	108.0	3.60	4.54	2300	2500	8.00
	120.0	105.0	3.50	4.41	2250	2500	10.00

MEMBRANE SCREENING DATA
SALT WATER FLUX DATA OPERATING PRESSURE 100 psi
BRACKISH WATER   TEMPERATURE 20C

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
CA-UF	15.0	136.0	9.07	11.42	2090	2300	9.13
	30.0	134.0	8.93	11.26	2070	2300	10.00
	45.0	133.0	8.87	11.17	2060	2290	10.04

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
PS-E500	15.5	180.0	11.61	14.63	2060	2410	14.52
	30.5	170.0	11.33	14.28	2080	2440	14.75
	45.5	145.0	9.67	12.18	2010	2300	12.61

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
CELLULOSE	15.5	2.4	0.15	0.20	1960	2410	18.67
13	30.5	2.4	0.16	0.20	1890	2440	22.54
	45.5	2.2	0.15	0.18	1950	2300	15.22

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
CELLULOSE	15.0	4.9	0.32	0.41	2100	2250	6.67
12	30.0	4.9	0.32	0.41	2180	2250	3.11
	45.0	4.7	0.31	0.39	2160	2275	5.05
	60.0	4.8	0.32	0.40	2160	2280	5.26

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
CELLULOSE	15.0	27.0	1.80	2.27	2100	2250	6.67
23	30.0	19.0	1.27	1.60	2180	2250	3.11
	45.0	6.0	0.40	0.50	2160	2275	5.05
	60.0	6.0	0.40	0.50	2160	2280	5.26

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
C.H.P.	15.0	191.0	12.73	16.04	1740	2300	24.35
Cat	30.0	183.0	12.20	15.37	1730	2300	24.78
	45.0	177.0	11.80	14.87	1720	2300	25.22

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
C.H.P.	15.0	139.0	9.27	11.68	1625	2300	29.35
B2	30.0	138.0	9.20	11.59	1610	2300	30.00
	45.0	136.0	9.07	11.42	1620	2300	29.57

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
C.H.P.	15.0	151.0	10.07	12.68	1700	2300	26.09
B3	30.0	149.0	9.93	12.52	1690	2300	26.52
	45.0	143.0	9.53	12.01	1675	2300	27.17

MEMBRANE SCREENING DATA
SALT WATER FLUX DATA
BRACKISH WATER TEMPERATURE 20 C

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
TFC G5	30.0	6.6	0.22	0.28	649	2200	70.50
	60.0	6.7	0.22	0.28	632	2210	71.40
	90.0	6.7	0.22	0.28	610	2210	72.40

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
TFC G10	30.0	23.0	0.77	0.97	610	2200	72.27
	60.0	22.8	0.76	0.96	600	2210	72.85
	90.0	22.5	0.75	0.95	590	2210	73.30

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
TFC G20	30.0	64.5	2.15	2.71	1040	2100	50.48
	60.0	63.0	2.10	2.65	1095	2180	49.77
	90.0	63.5	2.12	2.67	1100	2240	50.89

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
TFC G50	30.0	26.0	0.87	1.09	482	2100	77.05
	60.0	24.0	0.80	1.01	672	2180	69.17
	90.0	23.5	0.78	0.99	632	2240	71.79

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
CA-0K	15.0	77.0	5.13	6.47	2150	2150	0.00
	30.0	76.5	5.10	6.43	2140	2200	2.73
	45.5	78.0	5.03	6.34	2050	2100	2.38

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
CA-20K	15.0	67.5	4.50	5.67	2150	2150	0.00
	30.0	68.0	4.53	5.71	2200	2200	0.00
	45.5	69.0	4.45	5.61	2100	2100	0.00

				FLUX		CONDUCTIVITY	
MEMBRANE	TIME	ml	ml/min	M3/M2/D	PERM	FEED	% REJ
CA-50K	15.5	19.8	1.28	1.61	1145	2240	48.88
	30.5	19.0	1.27	1.60	1145	2210	48.19
	45.5	19.0	1.27	1.60	1140	2220	48.65

MEMBRANE PRESCREENING TEST DATA
TEMP18.5C pH 8.0 100 psi
FEED EMULSION (TDS 460 mg/L)

MEMBRANE	TIME MIN	PERM MLS	PERM MLS/MIN	FLUX RATE M3/M2/DAY	FEED TOC mg/L	PERM TOC mg/L	PERC REJ TOC	PERM COND uHmos/cm	PERM pH	TDS mg/L
TFC G50	30	28.0	0.93	1.01	1450	22.0	98.48			
	60	26.5	0.88	0.95	1845	86.0	95.34			
	90	25.5	0.85	0.92	1655	22.0	98.67			
	120	24.5	0.82	0.88	1445	22.0	98.48	245	7.82	248
	150	23.8	0.79	0.86	1655	22.0	98.67			
	180	23.0	0.77	0.83	1655	18.0	98.91			
	210	22.0	0.73	0.79	1655	22.0	98.67			
CA OK	30	92.0	3.07	3.31	1450	75.0	94.83			
	60	72.0	2.40	2.59	1845	8.0	99.57			
	90	52.0	1.73	1.87	1655	71.0	95.71			
	120	41.0	1.37	1.48	1445	65.0	95.50	760	8.11	352
	150	34.0	1.13	1.22	1655	67.0	95.95			
	180	29.0	0.97	1.04	1655	65.0	96.07			
	210	26.0	0.87	0.94	1655	73.0	95.59			
TFC G10	30	18.0	0.60	0.65	1450	15.0	98.97			
	60	18.0	0.60	0.65	1845	11.0	99.40			
	90	17.8	0.59	0.64	1655	8.0	99.52			
	120	17.5	0.58	0.63	1445	9.0	99.38	245	7.82	248
	150	17.5	0.58	0.63	1655	9.5	99.43			
	180	17.5	0.58	0.63	1655	8.5	99.49			
	210	17.0	0.57	0.61	1655	13.5	99.18			
TFC G20	30	34.0	1.13	1.22	1450	36.0	97.52			
	60	25.0	0.83	0.90	1845	38.0	97.94			
	90	21.7	0.72	0.78	1655	33.0	98.01			
	120	20.0	0.67	0.72	1445	33.0	97.72	460	7.6	196
	150	18.5	0.62	0.67	1655	29.0	98.25			
	180	17.0	0.57	0.61	1655	32.0	98.07			
	210	16.0	0.53	0.58	1655	32.0	98.07			
CA 20K	30	31.5	1.05	1.13	1450	121.0	91.66			
	60	19.0	0.63	0.68	1845	111.0	93.98			
	90	16.0	0.53	0.58	1655	103.0	93.78			
	120	14.0	0.47	0.50	1445	83.0	94.26	770	8.04	464
	150	13.0	0.43	0.47	1655	105.0	93.66			
	180	12.0	0.40	0.43	1655	96.0	94.20			
	210	11.0	0.37	0.40	1655	98.0	94.08			
CA 50K	30	20.0	0.67	0.72	1450	121.0	91.66			
	60	16.0	0.53	0.58	1845	111.0	93.98			
	90	14.0	0.47	0.50	1655	103.0	93.78			
	120	12.0	0.40	0.43	1445	83.0	94.26	435	7.92	108
	150	11.5	0.38	0.41	1655	105.0	93.66			
	180	10.0	0.33	0.36	1655	96.0	94.20			
	210	9.5	0.32	0.34	1655	98.0	94.08			

MEMBRANE PRESCREENING TEST DATA
TEMP 20C,pH 8.0,100 PSI
FEED EMULSION (TDS 700 mg/L)

RUN 2 MEMBRANE	TIME MIN	PERM MLS	PERM MLS/MIN	FLUX RATE M3/M2/DAY	FEED TOC mg/L	PERM TOC mg/L	PERC REJ TOC	PERM COND uHmos/cm	PERM pH	TDS mg/L
ZENON CHP	30	59.0	1.97	2.12	2255	44.0	98.05			
Cat	60	44.0	1.47	1.58	2195	64.5	97.06			
	90	35.0	1.17	1.26	2191	42.0	98.08			
	120	33.0	1.10	1.19	1881	49.0	97.40	535	7.96	NA
	150	34.0	1.13	1.22	1793	54.0	96.99			
	180	28.9	0.96	1.04	1342	40.0	97.02			
	210	25.5	0.85	0.92	1609	36.0	97.76			
ZENON CHP	30	92.0	3.07	3.31	2255	42.0	98.14			
B3	60	56.5	1.88	2.03	2195	60.0	97.27			
	90	40.9	1.36	1.47	2191	45.0	97.95			
	120	36.5	1.22	1.31	1881	45.0	97.61	520	8.01	312
	150	30.0	1.00	1.08	1793	37.0	97.94			
	180	21.0	0.70	0.76	1342	45.0	96.65			
	210	15.5	0.52	0.56	1609	36.0	97.76			
ZENON CHP	30	66.0	2.20	2.38	2255	57.0	97.47			
B2	60	34.0	1.13	1.22	2195	82.0	96.26			
	90	28.0	0.93	1.01	2191	44.0	97.99			
	120	24.5	0.82	0.88	1881	33.0	98.25	525	7.96	192
	150	21.8	0.73	0.78	1793	51.0	97.16			
	180	18.0	0.60	0.65	1342	36.0	97.32			
	210	14.8	0.49	0.53	1609	36.0	97.76			
PS E500	30	124.0	4.13	4.46	2255	73.0	96.76			
	60	59.0	1.97	2.12	2195	61.0	97.22			
	90	34.0	1.13	1.22	2191	56.0	97.44			
	120	34.0	1.13	1.22	1881	61.0	96.76	695	8.07	316
	150	29.0	0.97	1.04	1793	53.0	97.04			
	180	18.0	0.60	0.65	1342	55.0	95.90			
	210	13.0	0.43	0.47	1609	54.0	96.64			
PS 20K	30	41.0	1.37	1.48	2255	135.0	94.01			
	60	24.0	0.80	0.86	2195	109.0	95.03			
	90	20.0	0.67	0.72	2191	117.0	94.66			
	120	21.0	0.70	0.76	1881	108.0	94.26	800	8.11	564
	150	24.5	0.82	0.88	1793	113.0	93.70			
	180	21.0	0.70	0.76	1342	86.0	93.59			
	210	18.8	0.63	0.68	1609	79.0	95.09			
PS 50K	30	206.0	6.87	7.42	2255	56.0	97.52			
	60	110.0	3.67	3.96	2195	59.0	97.31			
	90	74.0	2.47	2.66	2191	65.0	97.03			
	120	69.0	2.30	2.48	1881	54.0	97.13	700	8	444
	150	56.0	1.87	2.02	1793	59.0	96.71			
	180	30.0	1.00	1.08	1342	56.0	95.83			
	210	21.0	0.70	0.76	1609	55.0	96.58			

MEMBRANE PRESCREENING TEST DATA
TEMP 20C,pH 8.0,100 psi
FEED EMULSION (TDS 780 mg/L)

RUN 3 MEMBRANE	TIME MIN	PERM MLS	PERM MLS/MIN	FLUX RATE M3/M2/DAY	FEED TOC mg/L	PERM TOC mg/L	PERC REJ TOC	PERM COND uHmos/cm	PERM pH	TDS mg/L
CELLULOSE 12	30	7.5	0.25	0.27	2255	292.0	87.05			
	60	6.0	0.20	0.22	2195	188.0	91.44			
	90	4.8	0.16	0.17	2191	159.0	92.74			
	120	4.5	0.15	0.16	1881	132.0	92.98	700	8.28	NA
	150	4.0	0.13	0.14	1793	131.0	92.69			
	180	3.0	0.10	0.11	1342	102.0	92.40			
	210	3.0	0.10	0.11	1609	110.0	93.16			
CELLULOSE 13	30	3.5	0.12	0.13	2255	585.0	74.06			
	60	3.0	0.10	0.11	2195	323.0	85.28			
	90	2.5	0.08	0.09	2191	176.0	91.97			
	120	2.8	0.09	0.10	1881	132.0	92.98	NA	NA	NA
	150	2.5	0.08	0.09	1793	101.0	94.37			
	180	2.5	0.08	0.09	1342	145.0	89.20			
	210	2.5	0.08	0.09	1609	154.0	90.43			
CELLULOSE 23	30	16.0	0.53	0.58	2255	204.0	90.95			
	60	13.3	0.44	0.48	2195	86.0	96.08			
	90	12.8	0.43	0.46	2191	74.0	96.62			
	120	12.5	0.42	0.45	1881	80.0	95.75	695	8.05	332
	150	12.0	0.40	0.43	1793	74.0	95.87			
	180	10.5	0.35	0.38	1342	69.0	94.86			
	210	10.5	0.35	0.38	1609	109.0	93.23			

MEMBRANE PORE SIZE EVALUATION
POLYETHYLENE GLYCOL 6000 M.W.
SCREENING TEST

AVERAGE FEED CONC. TOC 220 mg/L

CELLULOSE

TIME IN (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE M3/M2 D	TOC mg/L	% REJ TOC
15	3.0	0.20	0.22	48.5	77.95
30	3.5	0.23	0.25	10.0	95.45
60	5.5	0.18	0.20	18.0	91.82

POLYVINYLDIFLUORIDE

TIME IN (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE M3/M2 D	TOC mg/L	% REJ TOC
15	156.0	10.40	11.23	320.0	0.00
30	93.0	6.20	6.70	210.0	4.55
60	140.0	4.67	5.04	180.0	18.18

THIN FILM COMPOSITE

TIME IN (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE M3/M2 D	TOC mg/L	% REJ TOC
15	10.00	0.67	0.72	40.0	81.82
30	9.00	0.60	0.65	40.5	81.59
60	14.50	0.48	0.52	39.0	82.27

POLYSULFONE

TIME IN (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE M3/M2 D	TOC mg/L	% REJ TOC
15	475.00	31.67	34.20	230.0	0.00
30	350.00	23.33	25.20	210.0	4.55
60	509.00	16.97	18.32	180.0	18.18

CELLULOSE ACETATE

TIME IN (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE M3/M2 D	TOC mg/L	% REJ TOC
15	120.00	8.00	8.64	270.0	0.00
30	109.00	7.27	7.85	NA	0.00
60	194.00	6.47	6.98	200.0	9.09

CHP

TIME IN (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE M3/M2 D	TOC mg/L	% REJ TOC
15	149.50	9.97	10.76	200.0	9.09
30	138.00	9.20	9.94	150.0	31.82
60	240.00	8.00	8.64	220.0	0.00

APPENDIX D

RESULTS OF SCREENING TESTS

FLUX RATES FOR EMULSION SCREENING TEST NO. 1

Temp.=25°C pH = as is (7.5 - 8.0) Pressure = 100 - 150 psi

TIME (MIN)	CUMULATIVE VOLUME REDUCTION (%)	CHP FLUX RATE M3/M2·D	PS FLUX RATE M3/M2·D	PVDF FLUX RATE M3/M2·D	CA FLUX RATE M3/M2·D	TFC FLUX RATE M3/M2·D	CELLULOSE FLUX RATE M3/M2·D
60.00	0.00						
75.00	0.54	0.96	1.92	1.71	1.81	1.04	0.27
90.00	1.05	0.91	1.77	1.63	1.62	1.08	0.26
105.00	1.56	0.91	1.77	1.55	1.58	1.08	0.26
120.00	2.03	0.86	1.63	1.42	1.43	1.04	0.26
150.00	2.93	0.93	1.51	1.34	1.33	1.03	0.26
180.00	3.80	0.91	1.44	1.26	1.25	1.04	0.26
210.00	4.62	0.86	1.36	1.16	1.16	1.01	0.26
244.00	5.51	0.84	1.29	1.09	1.09	1.00	0.26
300.00	6.74	0.80	1.06	0.90	0.89	0.81	0.22
360.00	7.88	0.66	0.91	0.78	0.78	0.72	0.20
420.00	9.08	0.69	0.93	0.79	0.80	0.78	0.22
480.00	10.23	0.66	0.89	0.75	0.76	0.77	0.23
540.00	11.36	0.66	0.86	0.74	0.74	0.77	0.23
600.00	12.47	0.63	0.84	0.71	0.71	0.76	0.23
660.00	13.56	0.61	0.84	0.68	0.71	0.77	0.24
720.00	14.58	0.57	0.77	0.65	0.65	0.70	0.23
780.00	15.60	0.59	0.77	0.65	0.65	0.70	0.23
1260.00	22.64	0.52	0.65	0.56	0.56	0.59	0.22
1320.00	23.45	0.44	0.60	0.51	0.51	0.55	0.22
1380.00	24.27	0.45	0.61	0.52	0.52	0.56	0.23
1440.00	25.07	0.45	0.61	0.50	0.52	0.52	0.22
1500.00	25.86	0.44	0.59	0.50	0.50	0.53	0.23

ANALYTICAL DATA FOR SCREENING TEST NO.1

PARAMETER	UNITS	INIT FEED	FINAL FEED	MEAN FEED	C PERM	C %REJ	CA PERM	CA %REJ	PS PERM	PS %REJ	CHP PERM	CHP %REJ	TFC PERM	TFC %REJ	PVDF PERM	PVDF %REJ
pH		8.00	8.65	8.33	8.20		7.99		7.86		7.98		6.90		7.73	
COND	uHmos/cm	975	1240	1108	960	13.3	518	53.2	385	65.2	358	67.7	221	80.0	478	56.8
TURB	NTU	930	1250	1090	0.98	99.9	0.42	100.0	0.45	100.0	0.49	100.0	0.41	100.0	0.42	100.0
TOC	mg/L	1940	1740	1840	79.0	95.7	59.5	96.8	57.5	96.9	49.0	97.3	11.0	99.4	53	97.1
TDS	mg/L	966	1333	1150	644	44.0	308	73.2	562	51.1	362	68.5	176	84.7	366	68.2
HARDNESS	mg/L	4.56	NA	4.56	0.62	86.4	<0.15	100.0	<0.15	100.0	<0.15	100.0	<0.15	100.0	0.75	83.6
DIS SIO2	mg/L	67.5	NA	67.5	46.2	31.6	49.2	27.1	47.0	30.4	47.7	29.3	43.0	36.3	44.9	33.5
TSS	mg/L	1413	1280	1347	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA

RUN NO.1
% CUMULATIVE VOLUME REDUCTION
AUGUST 21 1984
VOLUME OF FEED 17100.00 MLS

TIME (MIN)	CHP PERM VOL (MLS)	PS PERM VOL (MLS)	PVDF PERM VOL (MLS)	CA PERM VOL (MLS)	TFC PERM VOL (MLS)	CELLULOSE PERM VOL (MLS)	TOTAL PERM VOL (MLS)	FEED pH	FEED TEMP (C)	CUM VOL RED (%)	FLOW TO CELLS (L/MIN)
15.00	21.00	40.00	49.00	50.50	15.00	4.00	0.00			0.00	1.92
30.00	20.50	28.00	31.00	34.00	15.50	4.00	0.00			0.00	1.92
45.00	12.00	24.50	25.00	28.50	15.00	3.40	0.00			0.00	1.92
60.00	10.00	21.00	22.00	25.00	14.00	3.50	0.00			0.00	1.92
75.00	10.00	20.00	21.00	23.50	14.50	3.70	92.70	8.00	23.00	0.54	1.92
90.00	9.50	18.50	20.00	21.00	15.00	3.60	87.60			1.05	1.92
105.00	9.50	18.50	19.00	20.50	15.00	3.60	86.10			1.56	1.84
120.00	9.00	17.00	17.50	18.50	14.50	3.60	80.10			2.03	1.88
150.00	19.50	31.50	33.00	34.50	28.50	7.10	154.10			2.93	1.80
180.00	19.00	30.00	31.00	32.50	29.00	7.30	148.80			3.80	2.06
210.00	18.00	28.50	28.50	30.00	28.00	7.20	140.20			4.62	2.00
244.00	20.00	30.50	30.50	32.00	31.50	8.30	152.80			5.51	2.00
300.00	31.00	41.50	41.50	43.00	42.00	11.20	210.20			6.74	1.94
360.00	27.50	38.00	38.50	40.50	40.00	10.90	195.40			7.88	2.04
420.00	29.00	39.00	39.00	41.50	43.50	12.30	204.30			9.08	2.00
480.00	27.50	37.00	37.00	39.50	43.00	12.60	196.60			10.23	1.94
540.00	27.50	36.00	36.50	38.50	43.00	12.90	194.40			11.36	1.92
600.00	26.50	35.00	35.00	37.00	42.00	13.00	188.50			12.47	1.84
660.00	25.50	35.00	33.50	37.00	42.50	13.50	187.00			13.56	1.92
720.00	24.00	32.00	32.00	34.00	39.00	13.00	174.00			14.58	1.92
780.00	24.50	32.00	32.00	33.50	39.00	13.00	174.00			15.60	1.92
1260.00	173.00	218.00	219.00	233.00	263.00	99.00	1205.00			22.64	1.92
1320.00	18.50	25.00	25.00	26.50	30.50	12.00	137.50			23.45	1.92
1380.00	19.00	25.50	25.50	27.00	31.00	12.50	140.50			24.27	1.92
1440.00	19.00	25.50	24.50	27.00	29.00	12.00	137.00			25.07	1.92
1500.00	18.50	24.50	24.50	26.00	29.50	12.50	135.50	8.65	23.00	25.86	1.92

RUN NO. I
MEMBRANE: CHP
PRESSURE= 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
75.00	10.00	0.67	18.11	0.72	0.96	
90.00	9.50	0.63	17.21	0.68	0.91	
105.00	9.50	0.63	17.21	0.68	0.91	
120.00	9.00	0.60	16.30	0.65	0.86	65.17
150.00	19.50	0.65	17.66	0.70	0.93	
180.00	19.00	0.63	17.21	0.68	0.91	
210.00	18.00	0.60	16.30	0.65	0.86	
244.00	20.00	0.59	15.98	0.64	0.84	70.79
300.00	31.00	0.55	15.04	0.60	0.80	
360.00	27.50	0.46	12.45	0.50	0.66	70.79
420.00	29.00	0.48	13.13	0.52	0.69	
480.00	27.50	0.46	12.45	0.50	0.66	70.79
540.00	27.50	0.46	12.45	0.50	0.66	
600.00	26.50	0.44	12.00	0.48	0.63	69.39
660.00	25.50	0.43	11.55	0.46	0.61	
720.00	24.00	0.40	10.87	0.43	0.57	65.17
780.00	24.50	0.41	11.09	0.44	0.59	
1260.00	173.00	0.36	9.79	0.39	0.52	
1320.00	18.50	0.31	8.38	0.33	0.44	
1380.00	19.00	0.32	8.60	0.34	0.45	70.79
1440.00	19.00	0.32	8.60	0.34	0.45	
1500.00	18.50	0.31	8.38	0.33	0.44	67.98

RUN NO. I
MEMBRANE: PVDF
PRESSURE= 130.00 psi
PRESSURE CORRECTION FACTOR 1.13

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	133.50					
75.00	21.00	1.40	38.04	1.51	1.71	
90.00	20.00	1.33	36.23	1.44	1.63	
105.00	19.00	1.27	34.42	1.37	1.55	
120.00	17.50	1.17	31.70	1.26	1.42	70.79
150.00	33.00	1.10	29.89	1.19	1.34	
180.00	31.00	1.03	24.91	1.12	1.26	
210.00	28.50	0.95	24.45	1.03	1.16	
244.00	30.50	0.90	22.64	0.97	1.09	70.79
300.00	41.50	0.74	21.74	0.80	0.90	
360.00	38.50	0.64	20.83	0.69	0.78	70.79
420.00	39.00	0.65	19.47	0.70	0.79	
480.00	37.00	0.62	18.57	0.67	0.75	73.60
540.00	36.50	0.61	16.53	0.66	0.74	
600.00	35.00	0.58	15.85	0.63	0.71	70.79
660.00	33.50	0.56	15.17	0.60	0.68	
720.00	32.00	0.53	14.49	0.58	0.65	70.79
780.00	32.00	0.53	14.49	0.58	0.65	
1260.00	219.00	0.46	12.40	0.49	0.56	
1320.00	25.00	0.42	11.32	0.45	0.51	
1380.00	25.50	0.43	11.55	0.46	0.52	70.79
1440.00	24.50	0.41	11.09	0.44	0.50	
1500.00	24.50	0.41	11.09	0.44	0.50	23.00

RUN NO. I
MEMBRANE: PS
PRESSURE= 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	118.50					
75.00	20.00	1.33	36.23	1.44	1.92	
90.00	18.50	1.23	33.51	1.33	1.77	
105.00	18.50	1.23	33.51	1.33	1.77	
120.00	17.00	1.13	30.79	1.22	1.63	65.17
150.00	31.50	1.05	28.53	1.13	1.51	
180.00	30.00	1.00	27.17	1.08	1.44	
210.00	28.50	0.95	25.81	1.03	1.36	
244.00	30.50	0.90	24.37	0.97	1.29	65.17
300.00	41.50	0.74	20.13	0.80	1.06	
360.00	38.00	0.63	17.21	0.68	0.91	67.98
420.00	39.00	0.65	17.66	0.70	0.93	
480.00	37.00	0.62	16.75	0.67	0.89	67.98
540.00	36.00	0.60	16.30	0.65	0.86	
600.00	35.00	0.58	15.85	0.63	0.84	67.98
660.00	35.00	0.58	15.85	0.63	0.84	
720.00	32.00	0.53	14.49	0.58	0.77	67.98
780.00	32.00	0.53	14.49	0.58	0.77	
1260.00	218.00	0.45	12.34	0.49	0.65	
1320.00	25.00	0.42	11.32	0.45	0.60	
1380.00	25.50	0.43	11.55	0.46	0.61	70.79
1440.00	25.50	0.43	11.55	0.46	0.61	
1500.00	24.50	0.41	11.09	0.44	0.59	67.98

RUN NO. I
MEMBRANE: CA
PRESSURE= 140.00 psi
PRESSURE CORRECTION FACTOR 1.07

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	149.00					
75.00	23.50	1.57	42.57	1.69	1.81	
90.00	21.00	1.40	38.04	1.51	1.62	
105.00	20.50	1.37	37.13	1.48	1.58	
120.00	18.50	1.23	33.51	1.33	1.43	65.17
150.00	34.50	1.15	31.25	1.24	1.33	
180.00	32.50	1.08	29.43	1.17	1.25	
210.00	30.00	1.00	27.17	1.08	1.16	
244.00	32.00	0.94	25.57	1.02	1.09	65.17
300.00	43.00	0.77	20.86	0.83	0.89	
360.00	40.50	0.68	18.34	0.73	0.78	73.60
420.00	41.50	0.69	18.79	0.75	0.80	
480.00	39.50	0.66	17.89	0.71	0.76	65.17
540.00	38.50	0.64	17.43	0.69	0.74	
600.00	37.00	0.62	16.75	0.67	0.71	73.60
660.00	37.00	0.62	16.75	0.67	0.71	
720.00	34.00	0.57	15.40	0.61	0.65	56.74
780.00	33.50	0.56	15.17	0.60	0.65	
1260.00	233.00	0.49	13.19	0.52	0.56	
1320.00	26.50	0.44	12.00	0.48	0.51	
1380.00	27.00	0.45	12.23	0.49	0.52	70.79
1440.00	27.00	0.45	12.23	0.49	0.52	
1500.00	26.00	0.43	11.77	0.47	0.50	62.36

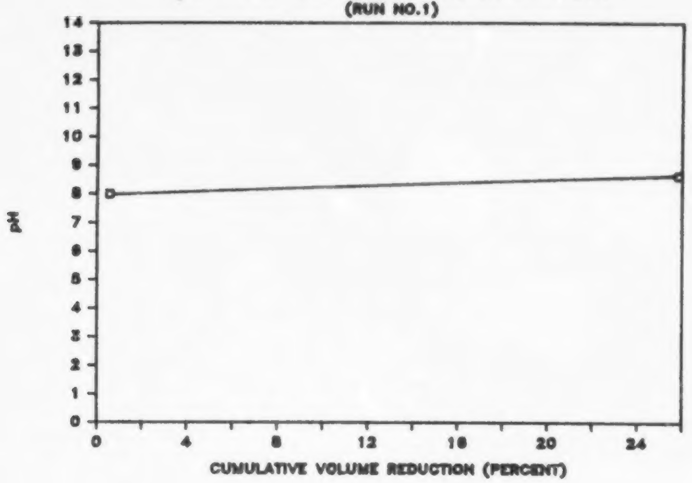
RUN NO. I
MEMBRANE: TFC
PRESSURE= 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	59.00					
75.00	14.50	0.97	26.26	1.04	1.04	
90.00	15.00	1.00	27.17	1.08	1.08	
105.00	15.00	1.00	27.17	1.08	1.08	
120.00	14.50	0.97	26.26	1.04	1.04	62.36
150.00	28.50	0.95	25.81	1.03	1.03	
180.00	29.00	0.97	26.26	1.04	1.04	
210.00	28.00	0.93	25.36	1.01	1.01	
244.00	31.50	0.93	25.17	1.00	1.00	65.17
300.00	42.00	0.75	20.38	0.81	0.81	
360.00	40.00	0.67	18.11	0.72	0.72	70.79
420.00	43.50	0.73	19.70	0.78	0.78	
480.00	43.00	0.72	19.47	0.77	0.77	73.60
540.00	43.00	0.72	19.47	0.77	0.77	
600.00	42.00	0.70	19.02	0.76	0.76	70.79
660.00	42.50	0.71	19.25	0.77	0.77	
720.00	39.00	0.65	17.66	0.70	0.70	76.41
780.00	39.00	0.65	17.66	0.70	0.70	
1260.00	263.00	0.55	14.89	0.59	0.59	
1320.00	30.50	0.51	13.81	0.55	0.55	
1380.00	31.00	0.52	14.04	0.56	0.56	70.79
1440.00	29.00	0.48	13.13	0.52	0.52	
1500.00	29.50	0.49	13.36	0.53	0.53	56.74

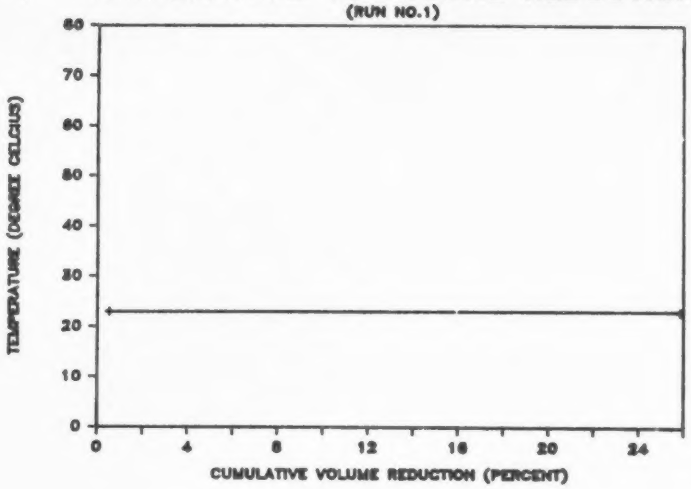
RUN NO. I
MEMBRANE: CELLULOSE
PRESSURE= 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	17.00					
75.00	3.70	0.25	6.70	0.27	0.27	
90.00	3.60	0.24	6.52	0.26	0.26	
105.00	3.60	0.24	6.52	0.26	0.26	
120.00	3.60	0.24	6.52	0.26	0.26	67.98
150.00	7.10	0.24	6.43	0.26	0.26	
180.00	7.30	0.24	6.61	0.26	0.26	
210.00	7.20	0.24	6.52	0.26	0.26	
244.00	8.30	0.24	6.63	0.26	0.26	67.98
300.00	11.20	0.20	5.43	0.22	0.22	
360.00	10.90	0.18	4.94	0.20	0.20	73.60
420.00	12.30	0.21	5.57	0.22	0.22	
480.00	12.60	0.21	5.71	0.23	0.23	67.98
540.00	12.90	0.22	5.84	0.23	0.23	
600.00	13.00	0.22	5.89	0.23	0.23	67.98
660.00	13.50	0.23	6.11	0.24	0.24	
720.00	13.00	0.22	5.89	0.23	0.23	67.98
780.00	13.00	0.22	5.89	0.23	0.23	
1260.00	99.00	0.21	5.60	0.22	0.22	
1320.00	12.00	0.20	5.43	0.22	0.22	
1380.00	12.50	0.21	5.66	0.23	0.23	70.79
1440.00	12.00	0.20	5.43	0.22	0.22	
1500.00	12.50	0.21	5.66	0.23	0.23	67.98

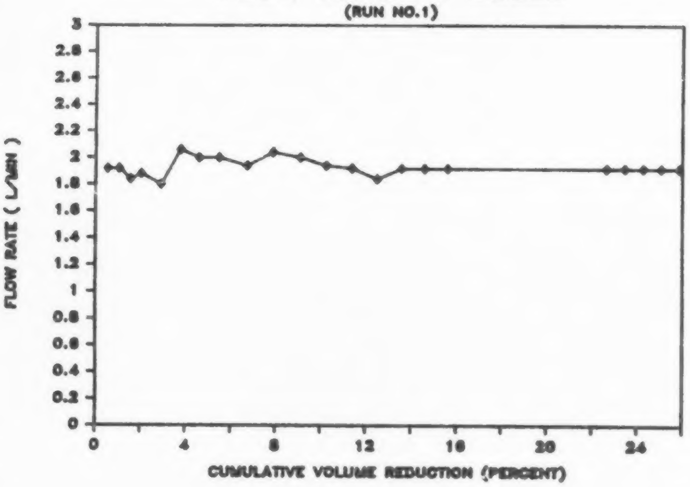
pH VS VOLUME REDUCTION
(RUN NO.1)



TEMPERATURE VS VOLUME REDUCTION
(RUN NO.1)



FLOW RATE TO CELLS
(RUN NO.1)



FLUX RATES FOR EMULSION SCREENING TEST NO. 2

Temp. = 45°C pH = as is (7.5 - 8.0) Pressure = 100 - 150 psi

TIME (MIN)	CUMULATIVE VOLUME REDUCTION (%)	CHP FLUX RATE M3/M2·D	PS FLUX RATE M3/M2·D	PVDF FLUX RATE M3/M2·D	CA FLUX RATE M3/M2·D	TFC FLUX RATE M3/M2·D	CELLULOSE FLUX RATE M3/M2·D
60.00	0.00						
90.00	1.28	2.01	1.99	1.87	1.89	1.12	0.32
120.00	2.34	1.63	1.60	1.55	1.54	1.04	0.29
150.00	3.43	1.63	1.63	1.51	1.52	1.17	0.32
180.00	4.41	1.41	1.44	1.34	1.35	1.12	0.31
210.00	5.25	1.22	1.24	1.12	1.17	0.99	0.27
240.00	6.07	1.15	1.17	1.10	1.12	0.97	0.27
300.00	7.62	1.11	1.10	1.02	1.04	0.95	0.28
360.00	9.13	1.08	1.05	0.98	1.02	0.95	0.28
420.00	10.59	1.07	1.02	0.94	0.96	0.94	0.27
480.00	11.96	0.97	0.96	0.87	0.90	0.88	0.29
540.00	13.28	0.90	0.90	0.83	0.87	0.87	0.28
600.00	14.53	0.86	0.86	0.79	0.82	0.82	0.27
660.00	15.73	0.84	0.83	0.75	0.78	0.78	0.26
1380.00	27.96	0.70	0.68	0.63	0.66	0.68	0.24
1440.00	29.14	0.81	0.79	0.71	0.71	0.80	0.32
1500.00	29.97	0.54	0.55	0.52	0.53	0.58	0.23

ANALYTICAL DATA FOR SCREENING TEST NO.2

PARAMETER	UNITS	INIT FEED	FINAL FEED	MEAN FEED	C PERM	C %REJ	CA PERM	CA %REJ	PS PERM	PS %REJ	CHP PERM	CHP %REJ	TFC PERM	TFC %REJ	PVDF PERM	PVDF %REJ
pH		7.56	8.40	7.98	8.12		7.86		8.02		7.82		7.79		7.73	
COND	uHmos/cm	900	1120	1010	850	15.8	470	53.5	485	52.0	470	53.5	195	80.7	420	58.4
TURB	NTU	1250	1400	1325	1.20	99.9	0.64	100.0	0.60	100.0	0.63	100.0	0.28	100.0	0.64	100.0
TOC	mg/L	2040	1990	2015	73.0	96.4	58.0	97.1	65.0	96.8	65.0	96.8	22.0	98.9	67.0	96.7
TDS	mg/L	980	NA	980	605	38.3	445	54.6	450	54.1	400	59.2	195	80.1	405	58.7
HARDNESS	mg/L	3.11	NA	3.11	1.37	55.9	<0.15	100.0	8.74	-181.0	<0.15	100.0	<0.15	100.0	<0.15	100.0
DIS SIO2	mg/L	69.6	NA	69.6	41.2	40.8	42.3	39.2	40.3	42.1	38.9	44.1	40.3	42.1	38.9	44.1
TSS	mg/L	1790	NA	1790	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA

RUN NO. 2
% CUMULATIVE VOLUME REDUCTION
AUGUST 23/84
VOLUME OF FEED 17100.00 MLS

TIME (MIN)	CHP PERM VOL (MLS)	PS PERM VOL (MLS)	PVDF PERM VOL (MLS)	CA PERM VOL (MLS)	TFC PERM VOL (MLS)	CELLULOSE PERM VOL (MLS)	TOTAL PERM VOL (MLS)	FEED pH	FEED TEMP (C)	CUM VOL RED (%)	FLOW TO CELLS (L/MIN)
15.00	39.00	36.50	43.00	49.00	12.50	3.50	0.00			0.00	2.00
30.00	31.00	31.50	35.00	40.00	16.50	4.60	0.00			0.00	2.20
45.00	25.00	26.00	29.00	31.50	15.00	4.50	0.00			0.00	1.92
60.00	23.00	24.50	26.50	28.50	15.00	4.40	0.00			0.00	1.94
90.00	42.00	41.50	46.00	49.00	31.00	8.80	218.30	7.56	42.00	1.28	2.04
120.00	34.00	33.50	38.00	40.00	29.00	8.00	182.50		45.00	2.34	1.92
150.00	34.00	34.00	37.00	39.50	32.50	8.80	185.80		43.00	3.43	1.92
180.00	29.50	30.00	33.00	35.00	31.00	8.50	167.00		45.00	4.41	1.84
210.00	25.50	26.00	27.50	30.50	27.50	7.60	144.60		44.00	5.25	1.88
240.00	24.00	24.50	27.00	29.00	27.00	7.60	139.10		43.00	6.07	1.80
300.00	46.50	46.00	50.00	54.00	53.00	15.50	265.00		45.00	7.62	2.06
360.00	45.00	44.00	48.00	53.00	53.00	15.50	258.50		46.00	9.13	2.00
420.00	44.50	42.50	46.00	50.00	52.00	15.20	250.20		45.00	10.59	2.00
480.00	40.50	40.00	43.00	46.50	49.00	16.00	235.00		44.00	11.96	1.94
540.00	37.50	37.50	41.00	45.00	48.50	15.70	225.20		45.00	13.28	2.04
600.00	36.00	36.00	39.00	42.50	45.50	15.00	214.00		44.00	14.53	2.00
660.00	35.00	34.50	37.00	40.50	43.50	14.50	205.00		44.00	15.73	1.94
1380.00	353.00	340.00	370.00	410.00	455.00	163.00	2091.00		52.00	27.96	1.92
1440.00	34.00	33.00	35.00	37.00	44.50	17.50	201.00		46.00	29.14	1.84
1500.00	22.50	23.00	25.50	27.50	32.00	12.50	143.00	8.40	42.00	29.97	1.92

RUN NO. II
MEMBRANE: CHP
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	42.00	1.40	38.04	1.51	2.01	72.20
120.00	34.00	1.13	30.79	1.22	1.63	
150.00	34.00	1.13	30.79	1.22	1.63	
180.00	29.50	0.98	26.72	1.06	1.41	62.36
210.00	25.50	0.85	23.09	0.92	1.22	
240.00	24.00	0.80	21.74	0.86	1.15	
300.00	46.50	0.78	21.06	0.84	1.11	70.79
360.00	45.00	0.75	20.38	0.81	1.08	
420.00	44.50	0.74	20.15	0.80	1.07	69.39
480.00	40.50	0.68	18.34	0.73	0.97	
540.00	37.50	0.63	16.98	0.68	0.90	69.39
600.00	36.00	0.60	16.30	0.65	0.86	
660.00	35.00	0.58	15.85	0.63	0.84	
1380.00	353.00	0.49	13.32	0.53	0.70	69.39
1440.00	34.00	0.57	15.40	0.61	0.81	72.20
1500.00	22.50	0.38	10.19	0.41	0.54	

RUN NO. II
MEMBRANE: PVDF
PRESSURE: 130.00 psi
PRESSURE CORRECTION FACTOR 1.13

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	46.00	1.53	41.66	1.66	1.87	
120.00	38.00	1.27	34.42	1.37	1.55	
150.00	37.00	1.23	33.51	1.33	1.51	
180.00	33.00	1.10	29.89	1.19	1.34	65.17
210.00	27.50	0.92	29.89	0.99	1.12	
240.00	27.00	0.90	24.91	0.97	1.10	
300.00	50.00	0.83	24.45	0.90	1.02	72.20
360.00	48.00	0.80	22.64	0.86	0.98	
420.00	46.00	0.77	21.74	0.83	0.94	70.79
480.00	43.00	0.72	20.83	0.77	0.87	
540.00	41.00	0.68	19.47	0.74	0.83	69.39
600.00	39.00	0.65	18.57	0.70	0.79	
660.00	37.00	0.62	16.75	0.67	0.75	
1380.00	370.00	0.51	13.96	0.56	0.63	72.20
1440.00	35.00	0.58	15.85	0.63	0.71	72.20
1500.00	25.50	0.43	11.55	0.46	0.52	

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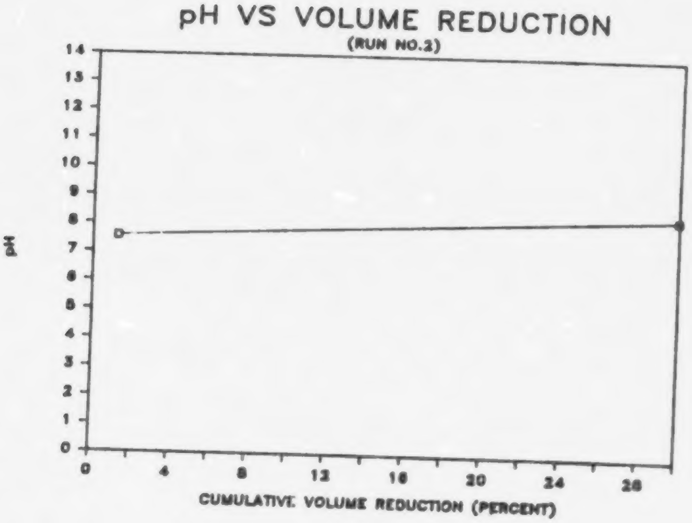
                          RUN NO.  II
                          MEMBRANE: PS
                          PRESSURE:  100.00 psi
                          PRESSURE CORRECTION FACTOR 1.33
   TIME    PERMEATE PERMEATE  FLUX      FLUX   CORRECTED SURFACE
            VOLUME   RATE     RATE      RATE   FLUX RATE TENSION
  (MIN)     (MLS)  (MLS/MIN)(US GFD) (M3/M2·D)(M3/M2·D)(DYNES/CM)
   60.00
   90.00    41.50    1.38    37.59     1.49     1.99    70.79
  120.00    33.50    1.12    30.34     1.21     1.60
  150.00    34.00    1.13    30.79     1.22     1.63
  180.00    30.00    1.00    27.17     1.08     1.44    63.77
  210.00    26.00    0.87    23.55     0.94     1.24
  240.00    24.50    0.82    22.19     0.88     1.17
  300.00    46.00    0.77    20.83     0.83     1.10    73.60
  360.00    44.00    0.73    19.92     0.79     1.05
  420.00    42.50    0.71    19.25     0.77     1.02    73.60
  480.00    40.00    0.67    18.11     0.72     0.96
  540.00    37.50    0.63    16.98     0.68     0.90    70.79
  600.00    36.00    0.60    16.30     0.65     0.86
  660.00    34.50    0.58    15.62     0.62     0.83
 1380.00   340.00    0.47    12.83     0.51     0.68    69.39
 1440.00    33.00    0.55    14.94     0.59     0.79    70.79
 1500.00    23.00    0.38    10.42     0.41     0.55
                                RUN NO.  II
                                MEMBRANE: CA
                                PRESSURE:  140.00 psi
                                PRESSURE CORRECTION FACTOR 1.07
   TIME    PERMEATE PERMEATE  FLUX      FLUX   CORRECTED SURFACE
            VOLUME   RATE     RATE      RATE   FLUX RATE TENSION
  (MIN)     (MLS)  (MLS/MIN)(US GFD) (M3/M2·D)(M3/M2·D)(DYNES/CM)
   60.00
   90.00    49.00    1.63    44.38     1.76     1.89    69.39
  120.00    40.00    1.33    36.23     1.44     1.54
  150.00    39.50    1.32    35.77     1.42     1.52
  180.00    35.00    1.17    31.70     1.26     1.35    62.36
  210.00    30.50    1.02    27.62     1.10     1.17
  240.00    29.00    0.97    26.26     1.04     1.12
  300.00    54.00    0.90    24.45     0.97     1.04    70.79
  360.00    53.00    0.88    24.00     0.95     1.02
  420.00    50.00    0.83    22.64     0.90     0.96    73.60
  480.00    46.50    0.78    21.06     0.84     0.90
  540.00    45.00    0.75    20.38     0.81     0.87    69.39
  600.00    42.50    0.71    19.25     0.77     0.82
  660.00    40.50    0.68    18.34     0.73     0.78
 1380.00   410.00    0.57    15.47     0.62     0.66    69.39
 1440.00    37.00    0.62    16.75     0.67     0.71    72.20
 1500.00    27.50    0.46    12.45     0.50     0.53
```

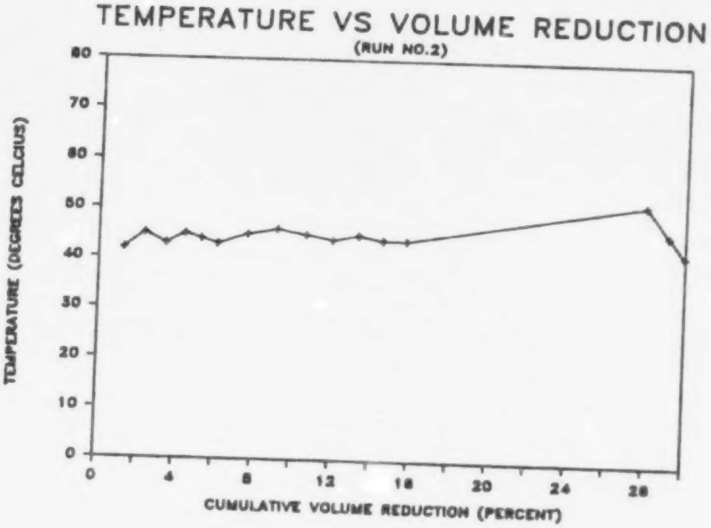
RUN NO. II
MEMBRANE: TFC
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

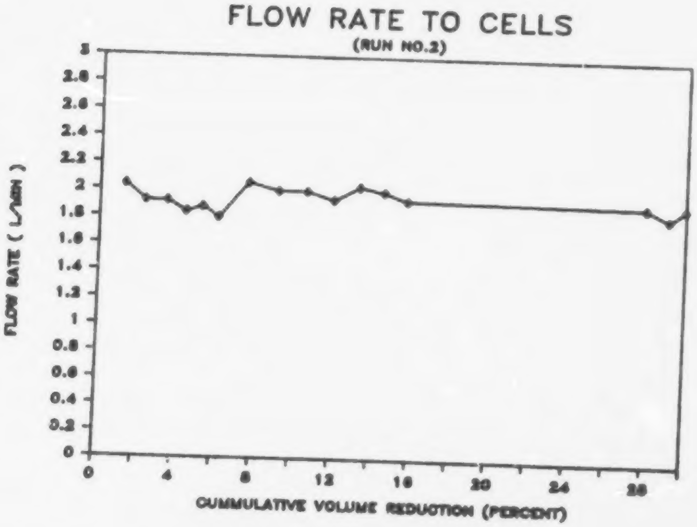
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	31.00	1.03	28.08	1.12	1.12	75.01
120.00	29.00	0.97	26.26	1.04	1.04	
150.00	32.50	1.08	29.43	1.17	1.17	
180.00	31.00	1.03	28.08	1.12	1.12	65.17
210.00	27.50	0.92	24.91	0.99	0.99	
240.00	27.00	0.90	24.45	0.97	0.97	
300.00	53.00	0.88	24.00	0.95	0.95	72.20
360.00	53.00	0.88	24.00	0.95	0.95	
420.00	52.00	0.87	23.55	0.94	0.94	70.79
480.00	49.00	0.82	22.19	0.88	0.88	
540.00	48.50	0.81	21.96	0.87	0.87	72.20
600.00	45.50	0.76	20.60	0.82	0.82	
660.00	43.50	0.73	19.70	0.78	0.78	
1380.00	455.00	0.63	17.17	0.68	0.68	70.79
1440.00	44.50	0.74	20.15	0.80	0.80	70.79
1500.00	32.00	0.53	14.49	0.58	0.58	

RUN NO. II
MEMBRANE: CELLULOSE
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	8.80	0.29	7.97	0.32	0.32	70.79
120.00	8.00	0.27	7.25	0.29	0.29	
150.00	8.80	0.29	7.97	0.32	0.32	
180.00	8.50	0.28	7.70	0.31	0.31	59.55
210.00	7.60	0.25	6.88	0.27	0.27	
240.00	7.60	0.25	6.88	0.27	0.27	
300.00	15.50	0.26	7.02	0.28	0.28	67.98
360.00	15.50	0.26	7.02	0.28	0.28	
420.00	15.20	0.25	6.88	0.27	0.27	70.79
480.00	16.00	0.27	7.25	0.29	0.29	
540.00	15.70	0.26	7.11	0.28	0.28	70.79
600.00	15.00	0.25	6.79	0.27	0.27	
660.00	14.50	0.24	6.57	0.26	0.26	
1380.00	163.00	0.23	6.15	0.24	0.24	67.98
1440.00	17.50	0.29	7.92	0.32	0.32	67.98
1500.00	12.50	0.21	5.66	0.23	0.23	


pH VS VOLUME REDUCTION
(RUN NO.2)
pH
0
1
2
3
4
5
6
7
8
9
10
11
12
13
14
0
4
8
12
16
20
24
28
CUMULATIVE VOLUME REDUCTION (PERCENT)


TEMPERATURE VS VOLUME REDUCTION
(RUN NO.2)
TEMPERATURE (DEGREES CELCIUS)
0
10
20
30
40
50
60
70
80
0
4
8
12
16
20
24
28
CUMULATIVE VOLUME REDUCTION (PERCENT)


FLOW RATE TO CELLS
(RUN NO.2)
FLOW RATE (L/MIN)
0
0.2
0.4
0.6
0.8
1
1.2
1.4
1.6
1.8
2
2.2
2.4
2.6
2.8
3
0
4
8
12
16
20
24
28
CUMMULATIVE VOLUME REDUCTION (PERCENT)

FLUX RATES FOR EMULSION SCREENING TEST NO. 3

Temp. =25 °C   pH = 5.5   Pressure = 100 - 150 psi

TIME (MIN)	CUMULATIVE VOLUME REDUCTION (%)	CHP FLUX RATE M3/M2·D	PS FLUX RATE M3/M2·D	PVDF FLUX RATE M3/M2·D	CA FLUX RATE M3/M2·D	TFC FLUX RATE M3/M2·D	CELLULOSE FLUX RATE M3/M2·D
60.00	0.00						
90.00	0.80	1.46	0.26	1.26	1.48	0.86	0.29
120.00	1.49	1.17	0.24	1.02	1.25	0.79	0.27
150.00	2.11	1.05	0.24	0.92	1.14	0.76	0.27
240.00	3.80	1.04	0.20	0.77	0.97	0.70	0.26
300.00	4.80	0.83	0.19	0.68	0.85	0.65	0.26
360.00	5.76	0.80	0.20	0.64	0.79	0.63	0.26
420.00	6.65	0.71	0.18	0.59	0.73	0.59	0.26
480.00	7.51	0.69	0.19	0.57	0.71	0.58	0.26
540.00	8.36	0.67	0.19	0.55	0.69	0.56	0.26
600.00	9.18	0.65	0.19	0.54	0.67	0.54	0.26
660.00	9.97	0.62	0.18	0.50	0.65	0.51	0.26
720.00	10.74	0.60	0.18	0.50	0.63	0.50	0.26
1260.00	16.84	0.56	0.16	0.42	0.54	0.43	0.24
1320.00	17.48	0.48	0.18	0.39	0.51	0.42	0.24
1380.00	18.13	0.50	0.18	0.40	0.51	0.41	0.26
1440.00	18.77	0.48	0.18	0.39	0.51	0.39	0.26

ANALYTICAL DATA FOR SCREENING TEST NO.3

PARAMETER	UNITS	INIT FEED	FINAL FEED	MEAN FEED	C PERM	C %REJ	CA PERM	CA %REJ	PS PERM	PS %REJ	CHP PERM	CHP %REJ	TFC PERM	TFC %REJ	PVDF PERM	PVDF %REJ
pH		5.50	6.60	6.05	6.70		6.65		6.01		6.68		6.78		6.65	
COND	uHmos/cm	875	950	913	1000		680	25.5	630	31.0	610	33.2	335	63.3	580	36.4
TURB	NTU	1150	1225	1188	0.84	99.9	0.44	100.0	2.20	99.8	0.40	100.0	0.28	100.0	0.29	100.0
TOC	mg/L	650	1400	1025	42.0	95.9	32.0	96.9	32.0	96.9	24.0	97.7	8.0	99.2	22.0	97.9
TDS	mg/L	1070	NA	1070	660	38.3	320	70.1	430	59.8	420	60.7	165	84.6	325	69.6
HARDNESS	mg/L	36.13	NA	36.13	27.69	23.4	7.15	80.2	3.45	90.5	4.45	87.7	0.75	97.9	4.2	88.4
DISS SIO2	mg/L	61.3	NA	61.3	40.0	34.7	40.9	33.3	35.7	41.8	42.3	31.0	42.3	31.0	41.2	32.8
TSS	mg/L	1706	NA	1706	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA

RUN NO.3
% CUMULATIVE VOLUME REDUCTION
AUGUST 27 1984 VOLUME OF FEED 17100.00 MLS

TIME (MIN)	CHP PERM VOL (MLS)	PS PERM VOL (MLS)	PVDF PERM VOL (MLS)	CA PERM VOL (MLS)	TFC PERM VOL (MLS)	CELLULOSE PERM VOL (MLS)	TOTAL PERM VOL (MLS)	FEED pH	FEED TEMP (C)	CUM VOL RED (%)	FLOW TO CELLS (L/MIN)
15.00	35.00	4.50	34.00	32.50	13.00	4.00	0.00			0.00	1.96
30.00	18.50	4.00	20.00	24.00	12.50	4.50	0.00			0.00	1.96
45.00	25.00	4.50	25.00	22.50	12.50	4.50	0.00			0.00	1.96
60.00	23.00	4.00	22.00	26.00	14.00	4.50	0.00	6.21	26.00	0.00	1.96
90.00	30.50	5.50	31.00	38.50	24.00	8.00	137.50		26.00	0.80	1.96
120.00	24.50	5.00	25.00	32.50	22.00	7.50	116.50	5.50	26.00	1.49	1.92
150.00	22.00	5.00	22.50	29.50	21.00	7.50	107.50		25.00	2.11	1.96
240.00	65.00	12.50	56.50	75.50	58.00	21.50	289.00	6.10	26.00	3.80	1.92
300.00	34.50	8.00	33.50	44.00	36.00	14.50	170.50	6.17	25.00	4.80	1.98
360.00	33.50	8.50	31.50	41.00	35.00	14.50	164.00	6.21	25.50	5.76	1.94
420.00	29.50	7.50	29.00	38.00	33.00	14.50	151.50	5.87	25.50	6.65	2.04
480.00	29.00	8.00	28.00	37.00	32.00	14.50	148.50	6.20	25.50	7.51	2.00
540.00	28.00	8.00	27.00	36.00	31.00	14.50	144.50	6.27	25.00	8.36	2.04
600.00	27.00	8.00	26.50	35.00	30.00	14.50	141.00		25.00	9.18	2.00
660.00	26.00	7.50	24.50	33.50	28.50	14.50	134.50		25.50	9.97	2.00
720.00	25.00	7.50	24.50	32.50	27.50	14.50	131.50		25.50	10.74	2.00
1260.00	209.00	60.00	184.00	252.00	216.00	122.00	1043.00	6.43	25.00	16.84	2.04
1320.00	20.00	7.50	19.00	26.50	23.50	13.50	110.00	6.49	25.00	17.48	2.04
1380.00	21.00	7.50	19.50	26.50	22.50	14.50	111.50	6.59	25.00	18.13	2.04
1440.00	20.00	7.50	19.00	26.50	21.50	14.50	109.00	6.64	25.00	18.77	2.04

RUN NO. III
MEMBRANE: CHP
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	30.50	1.02	27.62	1.10	1.46	66.58
120.00	24.50	0.82	22.19	0.88	1.17	
150.00	22.00	0.73	19.92	0.79	1.05	70.79
240.00	65.00	0.72	19.62	0.78	1.04	
300.00	34.50	0.58	15.62	0.62	0.83	70.79
360.00	33.50	0.56	15.17	0.60	0.80	
420.00	29.50	0.49	13.36	0.53	0.71	
480.00	29.00	0.48	13.13	0.52	0.69	73.60
540.00	28.00	0.47	12.68	0.50	0.67	
600.00	27.00	0.45	12.23	0.49	0.65	69.39
660.00	26.00	0.43	11.77	0.47	0.62	
720.00	25.00	0.42	11.32	0.45	0.60	
1260.00	209.00	0.39	10.52	0.42	0.56	
1320.00	20.00	0.33	9.06	0.36	0.48	67.98
1380.00	21.00	0.35	9.51	0.38	0.50	
1440.00	20.00	0.33	9.06	0.36	0.48	69.39

RUN NO. III
MEMBRANE: PVDF
PRESSURE: 130.00 psi
PRESSURE CORRECTION FACTOR 1.13

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	31.00	1.03	28.08	1.12	1.26	69.39
120.00	25.00	0.83	22.64	0.90	1.02	
150.00	22.50	0.75	20.38	0.81	0.92	67.98
240.00	56.50	0.63	17.06	0.68	0.77	
300.00	33.50	0.56	29.89	0.60	0.68	72.20
360.00	31.50	0.53	24.91	0.57	0.64	
420.00	29.00	0.48	24.45	0.52	0.59	
480.00	28.00	0.47	22.64	0.50	0.57	70.79
540.00	27.00	0.45	21.74	0.49	0.55	
600.00	26.50	0.44	20.83	0.48	0.54	70.79
660.00	24.50	0.41	19.47	0.44	0.50	
720.00	24.50	0.41	18.57	0.44	0.50	
1260.00	184.00	0.34	9.26	0.37	0.42	
1320.00	19.00	0.32	8.60	0.34	0.39	70.79
1380.00	19.50	0.33	8.83	0.35	0.40	
1440.00	19.00	0.32	8.60	0.34	0.39	70.79

RUN NO. III
MEMBRANE: PS
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	5.50	0.18	4.98	0.20	0.26	70.79
120.00	5.00	0.17	4.53	0.18	0.24	
150.00	5.00	0.17	4.53	0.18	0.24	69.39
240.00	12.50	0.14	3.77	0.15	0.20	
300.00	8.00	0.13	3.62	0.14	0.19	70.79
360.00	8.50	0.14	3.85	0.15	0.20	
420.00	7.50	0.13	3.40	0.14	0.18	
480.00	8.00	0.13	3.62	0.14	0.19	70.79
540.00	8.00	0.13	3.62	0.14	0.19	
600.00	8.00	0.13	3.62	0.14	0.19	70.79
660.00	7.50	0.13	3.40	0.14	0.18	
720.00	7.50	0.13	3.40	0.14	0.18	
1260.00	60.00	0.11	3.02	0.12	0.16	
1320.00	7.50	0.13	3.40	0.14	0.18	70.79
1380.00	7.50	0.13	3.40	0.14	0.18	
1440.00	7.50	0.13	3.40	0.14	0.18	67.98

RUN NO. III
MEMBRANE: CA
PRESSURE: 140.00 psi
PRESSURE CORRECTION FACTOR 1.07

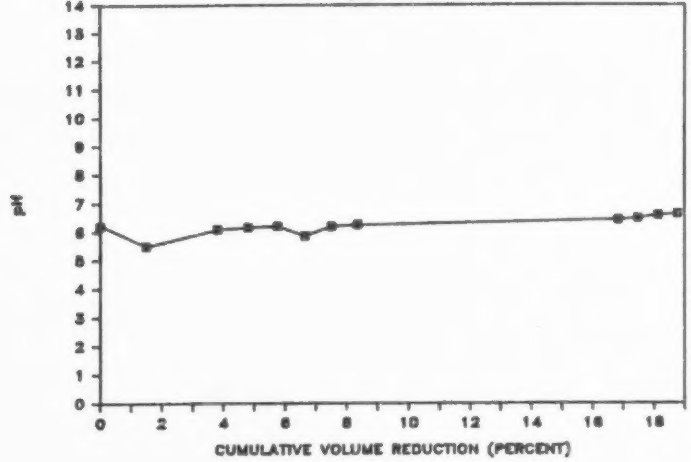
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	38.50	1.28	34.87	1.39	1.48	72.20
120.00	32.50	1.08	29.43	1.17	1.25	
150.00	29.50	0.98	26.72	1.06	1.14	63.77
240.00	75.50	0.84	22.79	0.91	0.97	
300.00	44.00	0.73	19.92	0.79	0.85	72.20
360.00	41.00	0.68	18.57	0.74	0.79	
420.00	38.00	0.63	17.21	0.68	0.73	
480.00	37.00	0.62	16.75	0.67	0.71	72.20
540.00	36.00	0.60	16.30	0.65	0.69	
600.00	35.00	0.58	15.85	0.63	0.67	72.20
660.00	33.50	0.56	15.17	0.60	0.65	
720.00	32.50	0.54	14.72	0.59	0.63	
1260.00	252.00	0.47	12.68	0.50	0.54	
1320.00	26.50	0.44	12.00	0.48	0.51	67.98
1380.00	26.50	0.44	12.00	0.48	0.51	
1440.00	26.50	0.44	12.00	0.48	0.51	67.98

RUN NO. III
MEMBRANE: TFC
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

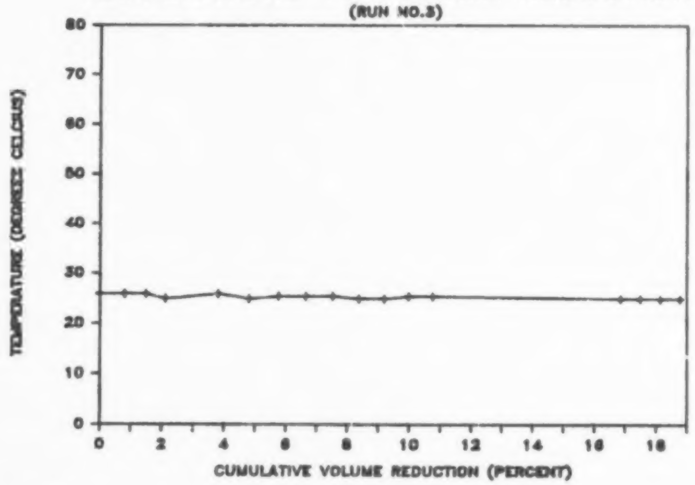
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	59.00					
90.00	24.00	0.80	21.74	0.86	0.86	67.98
120.00	22.00	0.73	19.92	0.79	0.79	
150.00	21.00	0.70	19.02	0.76	0.76	63.77
240.00	58.00	0.64	17.51	0.70	0.70	
300.00	36.00	0.60	16.30	0.65	0.65	73.60
360.00	35.00	0.58	15.85	0.63	0.63	
420.00	33.00	0.55	14.94	0.59	0.59	
480.00	32.00	0.53	14.49	0.58	0.58	70.79
540.00	31.00	0.52	14.04	0.56	0.56	
600.00	30.00	0.50	13.59	0.54	0.54	73.60
660.00	28.50	0.48	12.91	0.51	0.51	
720.00	27.50	0.46	12.45	0.50	0.50	
1260.00	216.00	0.40	10.87	0.43	0.43	
1320.00	23.50	0.39	10.64	0.42	0.42	70.79
1380.00	22.50	0.38	10.19	0.41	0.41	
1440.00	21.50	0.36	9.74	0.39	0.39	70.79

RUN NO. III
MEMBRANE: CELLULOSE
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

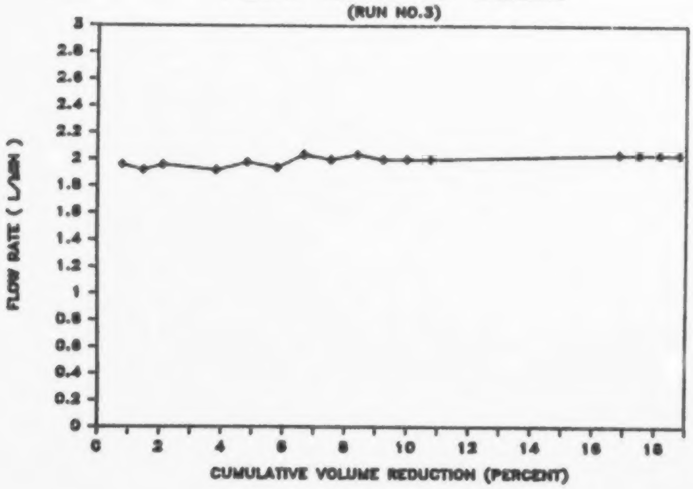
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	8.00	0.27	7.25	0.29	0.29	67.98
120.00	7.50	0.25	6.79	0.27	0.27	
150.00	7.50	0.25	6.79	0.27	0.27	62.36
240.00	21.50	0.24	6.49	0.26	0.26	
300.00	14.50	0.24	6.57	0.26	0.26	72.20
360.00	14.50	0.24	6.57	0.26	0.26	
420.00	14.50	0.24	6.57	0.26	0.26	
480.00	14.50	0.24	6.57	0.26	0.26	67.98
540.00	14.50	0.24	6.57	0.26	0.26	
600.00	14.50	0.24	6.57	0.26	0.26	67.98
660.00	14.50	0.24	6.57	0.26	0.26	
720.00	14.50	0.24	6.57	0.26	0.26	
1260.00	122.00	0.23	6.14	0.24	0.24	
1320.00	13.50	0.23	6.11	0.24	0.24	65.17
1380.00	14.50	0.24	6.57	0.26	0.26	
1440.00	14.50	0.24	6.57	0.26	0.26	66.58



pH VS VOLUME REDUCTION
(RUN NO.3)



TEMPERATURE VS VOLUME REDUCTION
(RUN NO.3)



FLOW RATE TO CELLS
(RUN NO.3)

FLUX RATES FOR EMULSION SCREENING TEST NO. 4

Temp. =45 °C pH = 5.5 Pressure = 100 - 150 psi

TIME (MIN)	CUMULATIVE VOLUME REDUCTION (%)	CHP FLUX RATE M3/M2·D	PS FLUX RATE M3/M2·D	PVDF FLUX RATE M3/M2·D	CA FLUX RATE M3/M2·D	TFC FLUX RATE M3/M2·D	CELLULOSE FLUX RATE M3/M2·D
60.00	0.00						
90.00	0.98	0.86	1.03	1.71	1.89	0.92	0.40
120.00	2.02	0.89	1.08	1.81	2.04	1.01	0.45
150.00	2.88	0.81	0.93	1.44	1.66	0.76	0.38
180.00	3.67	0.77	0.93	1.26	1.54	0.70	0.34
240.00	5.18	0.78	0.95	1.12	1.46	0.67	0.33
300.00	6.67	0.72	1.17	1.03	1.41	0.62	0.32
360.00	8.11	0.69	1.14	0.98	1.35	0.59	0.33
420.00	9.36	0.65	1.04	0.83	1.14	0.49	0.30
480.00	10.60	0.54	1.01	0.83	1.16	0.51	0.32
540.00	11.77	0.50	0.98	0.78	1.08	0.48	0.32
600.00	12.95	0.55	1.03	0.77	1.05	0.45	0.32
660.00	14.05	0.50	0.97	0.72	0.97	0.42	0.30
720.00	15.11	0.48	0.96	0.69	0.94	0.41	0.29
1380.00	24.91	0.42	0.79	0.58	0.81	0.29	0.27
1440.00	25.69	0.29	0.68	0.53	0.74	0.24	0.27
1500.00	26.37	0.28	0.60	0.47	0.64	0.20	0.23

ANALYTICAL DATA FOR SCREENING TEST NO.4

PARAMETER	UNITS	INIT FEED	FINAL FEED	MEAN FEED	C PERM	C %REJ	CA PERM	CA %REJ	PS PERM	PS %REJ	CHP PERM	CHP %REJ	TFC PERM	TFC %REJ	PVDF PERM	PVDF %REJ
pH		6.05	6.63	6.34	6.94		7.01		6.99		6.92		6.98		6.91	
COND	uHmos/cm	850	1150	1000	940	6.0	540	46.0	550	45.0	465	53.5	220	78.0	470	53.0
TURB	NTU	850	1800	1325	1.00	99.9	0.68	99.9	0.52	100.0	0.60	100.0	0.40	100.0	0.72	99.9
TOC	mg/L	1202	1320	1261	62.0	95.1	32.0	97.5	32.0	97.5	21.0	98.3	8.0	99.4	26.0	97.9
TDS	mg/L	1050	NA	1050	740	29.5	455	56.7	475	54.8	460	56.2	230	78.1	395	62.4
HARDNESS	mg/L	13.74	NA	13.74	22.85	-66.3	4.49	67.3	2.53	81.6	2.57	81.3	0.25	98.2	2.82	79.5
DIS SI02	mg/L	84.8	NA	84.8	68.7	19.0	69.8	17.7	69.8	17.7	68.2	19.6	58.8	30.7	71.6	15.6
TSS	mg/L	1015	NA	1015	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA

RUN NO.4
% CUMULATIVE VOLUME REDUCTION
AUGUST 29 1984
VOLUME OF FEED 17100.00 MLS

TIME (MIN)	CHP PERM VOL (MLS)	PS PERM VOL (MLS)	PVDF PERM VOL (MLS)	CA PERM VOL (MLS)	TFC PERM VOL (MLS)	CELLULOSE PERM VOL (MLS)	TOTAL PERM VOL (MLS)	FEED pH	FEED TEMP (C)	CUM VOL RED (%)	FLOW TO CELLS (L/MIN)
15.00	16.00	18.00	48.50	53.00	14.00	6.00	0.00			0.00	1.84
30.00	11.00	18.00	33.00	38.00	16.00	7.50	0.00			0.00	2.12
45.00	13.00	18.00	32.50	38.00	18.00	7.50	0.00			0.00	2.28
60.00	11.00	14.50	27.50	32.00	16.00	6.50	0.00	5.70	43.00	0.00	1.80
90.00	18.00	21.50	42.00	49.00	25.50	11.00	167.00	5.70	42.00	0.98	1.80
120.00	18.50	22.50	44.50	53.00	28.00	12.50	179.00	7.50	46.00	2.02	2.32
150.00	17.00	19.50	35.50	43.00	21.00	10.50	146.50	7.50	44.00	2.88	1.92
180.00	16.00	19.50	31.00	40.00	19.50	9.50	135.50	7.66	42.00	3.67	1.88
240.00	32.50	39.50	55.00	76.00	37.00	18.50	258.50	7.66	47.00	5.18	2.06
300.00	30.00	49.00	50.50	73.00	34.50	17.50	254.50	6.11	42.00	6.67	1.88
360.00	29.00	47.50	48.00	70.00	33.00	18.50	246.00	6.63	48.00	8.11	2.00
420.00	27.00	43.50	41.00	59.00	27.00	16.50	214.00	6.78	44.00	9.36	1.80
480.00	22.50	42.00	41.00	60.00	28.50	18.00	212.00	6.03	45.00	10.60	2.04
540.00	21.00	41.00	38.50	56.00	26.50	17.50	200.50	5.93	44.00	11.77	2.00
600.00	23.00	43.00	38.00	54.50	25.00	17.50	201.00	6.27	44.00	12.95	1.92
660.00	21.00	40.50	35.50	50.50	23.50	16.50	187.50	6.36	45.00	14.05	1.88
720.00	20.00	40.00	34.00	49.00	22.50	16.00	181.50	6.55	44.00	15.11	1.88
1380.00	193.00	365.00	315.00	462.00	175.00	166.00	1676.00	6.05	48.00	24.91	2.00
1440.00	12.00	28.50	26.00	38.50	13.50	15.00	133.50	6.54	46.00	25.69	2.16
1500.00	11.50	25.00	23.00	33.00	11.00	12.50	116.00	6.73	44.00	26.37	2.00

RUN NO. IV
MEMBRANE: CHP
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE MLS/MIN	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						73.60
90.00	18.00	0.60	16.30	0.65	0.86	
120.00	18.50	0.62	16.75	0.67	0.89	70.79
150.00	17.00	0.57	15.40	0.61	0.81	
180.00	16.00	0.53	14.49	0.58	0.77	
240.00	32.50	0.54	14.72	0.59	0.78	70.79
300.00	30.00	0.50	13.59	0.54	0.72	
360.00	29.00	0.48	13.13	0.52	0.69	70.79
420.00	27.00	0.45	12.23	0.49	0.65	
480.00	22.50	0.38	10.19	0.41	0.54	75.01
540.00	21.00	0.35	9.51	0.38	0.50	
600.00	23.00	0.38	10.42	0.41	0.55	67.98
660.00	21.00	0.35	9.51	0.38	0.50	
720.00	20.00	0.33	9.06	0.36	0.48	70.79
1380.00	193.00	0.29	7.95	0.32	0.42	
1440.00	12.00	0.20	5.43	0.22	0.29	
1500.00	11.50	0.19	5.21	0.21	0.28	69.39

RUN NO. IV
MEMBRANE: PVDF
PRESSURE: 120.00 psi
PRESSURE CORRECTION FACTOR 1.13

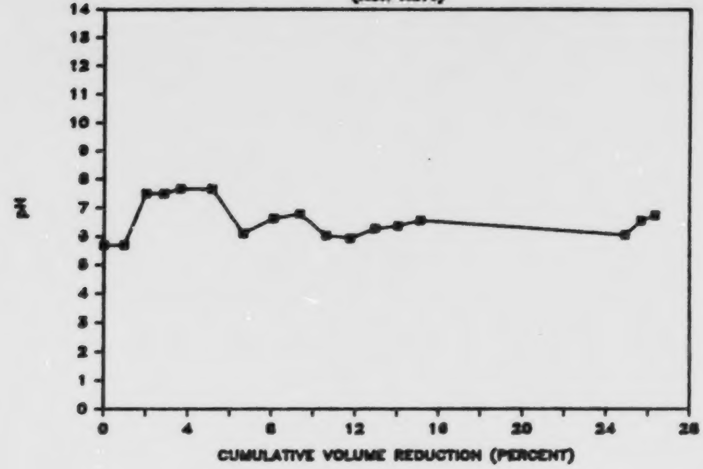
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE MLS/MIN	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	133.50					72.20
90.00	42.00	1.40	38.04	1.51	1.71	
120.00	44.50	1.48	40.30	1.60	1.81	66.58
150.00	35.50	1.18	32.15	1.28	1.44	
180.00	31.00	1.03	28.08	1.12	1.26	
240.00	55.00	0.92	29.89	0.99	1.12	70.79
300.00	50.50	0.84	24.91	0.91	1.03	
360.00	48.00	0.80	24.45	0.86	0.98	69.39
420.00	41.00	0.68	22.64	0.74	0.83	
480.00	41.00	0.68	21.74	0.74	0.83	73.60
540.00	38.50	0.64	20.83	0.69	0.78	
600.00	38.00	0.63	19.47	0.68	0.77	70.79
660.00	35.50	0.59	18.57	0.64	0.72	
720.00	34.00	0.57	15.40	0.61	0.69	72.20
1380.00	315.00	0.48	12.97	0.52	0.58	
1440.00	26.00	0.43	11.77	0.47	0.53	
1500.00	23.00	0.38	10.42	0.41	0.47	67.98

```
                      RUN NO.  IV
                      MEMBRANE: PS
                      PRESSURE:  100.00 psi
                      PRESSURE CORRECTION FACTOR 1.33
   TIME    PERMEATE PERMEATE   FLUX       FLUX      FLUX     SURFACE
            VOLUME    RATE     RATE       RATE      RATE     TENSION
   (MIN)    (MLS)   MLS/MIN (US GFD) (M3/M2·D)(M3/M2·D)(DYNES/CM)
    60.00   118.50                                            70.79
    90.00    21.50     0.72    19.47      0.77      1.03
   120.00    22.50     0.75    20.38      0.81      1.08      69.39
   150.00    19.50     0.65    17.66      0.70      0.93
   180.00    19.50     0.65    17.66      0.70      0.93
   240.00    39.50     0.66    17.89      0.71      0.95      73.60
   300.00    49.00     0.82    22.19      0.88      1.17
   360.00    47.50     0.79    21.51      0.86      1.14      70.79
   420.00    43.50     0.73    19.70      0.78      1.04
   480.00    42.00     0.70    19.02      0.76      1.01      70.79
   540.00    41.00     0.68    18.57      0.74      0.98
   600.00    43.00     0.72    19.47      0.77      1.03      67.98
   660.00    40.50     0.68    18.34      0.73      0.97
   720.00    40.00     0.67    18.11      0.72      0.96      70.79
  1380.00   365.00     0.55    15.03      0.60      0.79
  1440.00    28.50     0.48    12.91      0.51      0.68
  1500.00    25.00     0.42    11.32      0.45      0.60      70.79
                      RUN NO.  IV
                      MEMBRANE: CA
                      PRESSURE:  140.00 psi
                      PRESSURE CORRECTION FACTOR     1.07
   TIME    PERMEATE PERMEATE   FLUX       FLUX      FLUX     SURFACE
            VOLUME    RATE     RATE       RATE      RATE     TENSION
   (MIN)    (MLS)   MLS/MIN (US GFD) (M3/M2·D)(M3/M2·D)(DYNES/CM)
    60.00   149.00                                            72.20
    90.00    49.00     1.63    44.38      1.76      1.89
   120.00    53.00     1.77    48.00      1.91      2.04      70.79
   150.00    43.00     1.43    38.94      1.55      1.66
   180.00    40.00     1.33    36.23      1.44      1.54
   240.00    76.00     1.27    34.42      1.37      1.46      70.79
   300.00    73.00     1.22    33.06      1.31      1.41
   360.00    70.00     1.17    31.70      1.26      1.35      73.60
   420.00    59.00     0.98    26.72      1.06      1.14
   480.00    60.00     1.00    27.17      1.08      1.16      70.79
   540.00    56.00     0.93    25.36      1.01      1.08
   600.00    54.50     0.91    24.68      0.98      1.05      67.98
   660.00    50.50     0.84    22.87      0.91      0.97
   720.00    49.00     0.82    22.19      0.88      0.94      72.20
  1380.00   462.00     0.70    19.02      0.76      0.81
  1440.00    38.50     0.64    17.43      0.69      0.74
  1500.00    33.00     0.55    14.94      0.59      0.64      72.20
```

```
                        RUN NO.  IV
                        MEMBRANE: TFC
                        PRESSURE:  150.00 psi
                        PRESSURE CORRECTION FACTOR 1.00
  TIME    PERMEATE PERMEATE   FLUX     FLUX     FLUX     SURFACE
           VOLUME    RATE     RATE     RATE     RATE     TENSION
  (MIN)    (MLS)   MLS/MIN (US GFD) (M3/M2·D)(M3/M2·D)(DYNES/CM)
   60.00    59.00                                         70.79
   90.00    25.50     0.85    23.09     0.92     0.92
  120.00    28.00     0.93    25.36     1.01     1.01     70.79
  150.00    21.00     0.70    19.02     0.76     0.76
  180.00    19.50     0.65    17.66     0.70     0.70
  240.00    37.00     0.62    16.75     0.67     0.67     73.60
  300.00    34.50     0.58    15.62     0.62     0.62
  360.00    33.00     0.55    14.94     0.59     0.59     70.79
  420.00    27.00     0.45    12.23     0.49     0.49
  480.00    28.50     0.48    12.91     0.51     0.51     70.79
  540.00    26.50     0.44    12.00     0.48     0.48
  600.00    25.00     0.42    11.32     0.45     0.45     66.58
  660.00    23.50     0.39    10.64     0.42     0.42
  720.00    22.50     0.38    10.19     0.41     0.41     66.58
 1380.00   175.00     0.27     7.20     0.29     0.29
 1440.00    13.50     0.23     6.11     0.24     0.24
 1500.00    11.00     0.18     4.98     0.20     0.20     67.98
                        RUN NO.  IV
                        MEMBRANE: CELLULOSE
                        PRESSURE:  150.00 psi
                        PRESSURE CORRECTION FACTOR 1.00
  TIME    PERMEATE PERMEATE   FLUX     FLUX     FLUX     SURFACE
           VOLUME    RATE     RATE     RATE     RATE     TENSION
  (MIN)    (MLS)   MLS/MIN (US GFD) (M3/M2·D)(M3/M2·D)(DYNES/CM)
   60.00    17.00                                         67.98
   90.00    11.00     0.37     9.96     0.40     0.40
  120.00    12.50     0.42    11.32     0.45     0.45     67.98
  150.00    10.50     0.35     9.51     0.38     0.38
  180.00     9.50     0.32     8.60     0.34     0.34
  240.00    18.50     0.31     8.38     0.33     0.33     70.79
  300.00    17.50     0.29     7.92     0.32     0.32
  360.00    18.50     0.31     8.38     0.33     0.33     69.39
  420.00    16.50     0.28     7.47     0.30     0.30
  480.00    18.00     0.30     8.15     0.32     0.32     63.77
  540.00    17.50     0.29     7.92     0.32     0.32
  600.00    17.50     0.29     7.92     0.32     0.32     70.79
  660.00    16.50     0.28     7.47     0.30     0.30
  720.00    16.00     0.27     7.25     0.29     0.29     70.79
 1380.00   166.00     0.25     6.83     0.27     0.27
 1440.00    15.00     0.25     6.79     0.27     0.27
 1500.00    12.50     0.21     5.66     0.23     0.23     70.79
```

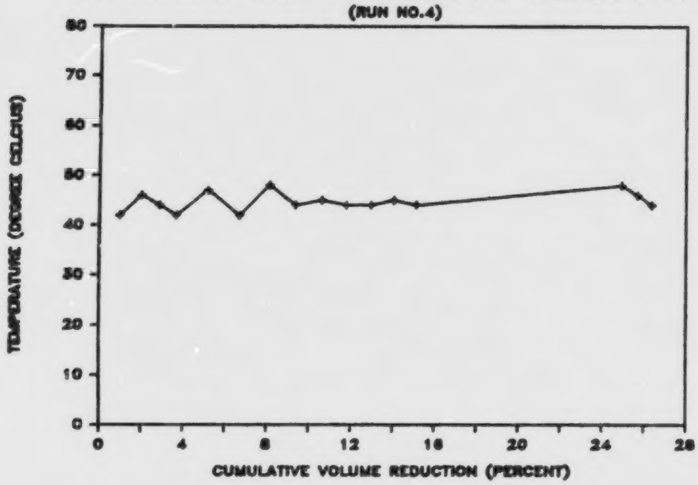
pH VS VOLUME REDUCTION

(RUN NO.4)



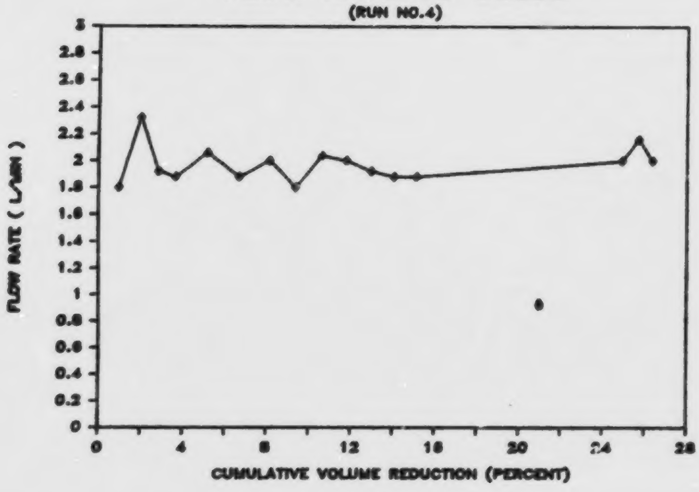
TEMPERATURE VS VOLUME REDUCTION

(RUN NO.4)



FLOW RATE TO CELLS

(RUN NO.4)



FLUX RATES FOR EMULSION SCREENING TEST NO. 5

Temp. = 25°C pH = 9.5 Pressure = 100 - 150 psi

TIME (MIN)	CUMULATIVE VOLUME REDUCTION (%)	CHP FLUX RATE M3/M2·D	PS FLUX RATE M3/M2·D	PVDF FLUX RATE M3/M2·D	CA FLUX RATE M3/M2·D	TFC FLUX RATE M3/M2·D	CELLULOSE FLUX RATE M3/M2·D
60.00	0.00						
90.00	1.28	1.68	1.92	1.65	2.29	1.30	0.29
120.00	2.25	1.22	1.44	1.22	1.75	1.01	0.23
150.00	3.05	1.01	1.17	1.00	1.44	0.85	0.22
180.00	3.81	0.96	1.10	0.94	1.35	0.85	0.22
210.00	4.56	0.93	1.05	0.89	1.31	0.85	0.22
240.00	5.27	0.89	1.01	0.85	1.23	0.83	0.22
300.00	6.62	0.87	0.93	0.78	1.18	0.81	0.21
360.00	7.99	0.84	0.95	0.75	1.19	0.85	0.23
420.00	9.31	0.84	0.87	0.71	1.18	0.81	0.24
480.00	10.55	0.77	0.81	0.66	1.12	0.76	0.24
540.00	11.72	0.71	0.75	0.61	1.08	0.71	0.24
600.00	12.75	0.63	0.67	0.55	0.91	0.61	0.22
660.00	13.73	0.61	0.65	0.53	0.88	0.59	0.22
720.00	14.65	0.56	0.62	0.49	0.82	0.54	0.21
1380.00	23.79	0.53	0.54	0.43	0.72	0.48	0.21
1440.00	24.56	0.47	0.51	0.40	0.66	0.45	0.21
1500.00	25.31	0.45	0.50	0.39	0.65	0.44	0.21

ANALYTICAL DATA FOR SCREENING TEST NO.5

PARAMETER	UNITS	INIT FEED	FINAL FEED	MEAN FEED	C PERM	C %REJ	CA PERM	CA %REJ	PS PERM	PS %REJ	CHP PERM	CHP %REJ	TFC PERM	TFC %REJ	PVDF PERM	PVDF %REJ
pH		9.62	9.35	9.49	9.27		9.14		8.95		9.01		8.62		8.93	
COND	uHmos/cm	1140	1280	1210	1040	14.0	800	33.9	610	49.6	640	47.1	274	77.4	585	51.7
TURB	NTU	950	1750	1350	0.87	99.9	0.76	99.9	0.44	100.0	0.47	100.0	0.12	100.0	0.38	100.0
TOC	mg/L	1650	1050	1350	92	93.2	88	93.5	66	95.1	76.5	94.3	44.5	96.7	53.5	96.0
TDS	mg/L	1155	NA	1155	705	39.0	550	52.4	420	63.6	445	61.5	160	86.1	430	62.8
HARDNESS	mg/L	6.06	NA	6.06	1.66	72.6	0.29	95.2	<0.15	100.0	<0.15	100.0	<0.15	100.0	<0.15	100.0
DIS SIO2	mg/L	82.5	NA	82.5	47	43.0	55.7	32.5	54.1	34.4	55.9	32.2	49.7	39.8	55.2	33.1
TSS	mg/L	1580	NA	1580	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA

RUN NO.5
% CUMULATIVE VOLUME REDUCTION
SEPTEMBER 4 1984 VOLUME OF FEED 17100.00 MLS

TIME (MIN)	CHP PERM VOL (MLS)	PS PERM VOL (MLS)	PVDF PERM VOL (MLS)	CA PERM VOL (MLS)	TFC PERM VOL (MLS)	CELLULOSE PERM VOL (MLS)	TOTAL PERM VOL (MLS)	FEED pH	FEED TEMP (C)	CUM VOL RED (%)	FLOW TO CELLS (L/MIN)
15.00	41.50	32.50	48.00	44.00	17.00	4.50	0.00			0.00	2.00
30.00	26.50	40.50	32.50	39.00	20.50	4.50	0.00			0.00	2.02
45.00	22.00	29.00	29.50	35.00	20.00	4.50	0.00			0.00	2.00
60.00	19.00	24.50	24.00	32.50	19.00	4.00	0.00	9.47	27.00	0.00	2.04
90.00	35.00	40.00	40.50	59.50	36.00	8.00	219.00	9.47	27.00	1.28	2.00
120.00	25.50	30.00	30.00	45.50	28.00	6.50	165.50	9.49	23.00	2.25	1.94
150.00	21.00	24.50	24.50	37.50	23.50	6.00	137.00	9.49	23.00	3.05	1.84
180.00	20.00	23.00	23.00	35.00	23.50	6.00	130.50	9.49	24.00	3.81	1.98
210.00	19.50	22.00	22.00	34.00	23.50	6.00	127.00	9.51	23.00	4.56	1.96
240.00	18.50	21.00	21.00	32.00	23.00	6.00	121.50	9.51	23.00	5.27	2.00
300.00	36.50	39.00	38.50	61.50	45.00	11.50	232.00	9.15	24.00	6.62	1.92
360.00	35.00	39.50	37.00	62.00	47.00	13.00	233.50	9.43	25.00	7.99	1.92
420.00	35.00	36.50	35.00	61.50	45.00	13.50	226.50	9.41	24.00	9.31	2.04
480.00	32.00	34.00	32.50	58.00	42.00	13.50	212.00	9.40	24.00	10.55	2.00
540.00	29.50	31.50	30.00	56.00	39.50	13.50	200.00	9.39	25.00	11.72	1.96
600.00	26.50	28.00	27.00	47.50	34.00	12.00	175.00	9.38	24.00	12.75	2.08
660.00	25.50	27.00	26.00	45.50	33.00	12.00	169.00	9.36	24.00	13.73	1.96
720.00	23.50	26.00	24.00	42.50	30.00	11.50	157.50	9.36	24.00	14.65	2.00
1380.00	245.00	250.00	235.00	410.00	295.00	127.00	1562.00	9.35	24.00	23.79	2.00
1440.00	19.50	21.50	19.50	34.50	25.00	11.50	131.50	9.35	23.00	24.56	1.96
1500.00	19.00	21.00	19.00	34.00	24.50	11.50	129.00	9.35	23.00	25.31	2.00

RUN NO. V
MEMBRANE: CHP
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	35.00	1.17	31.70	1.26	1.68	70.79
120.00	25.50	0.85	23.09	0.92	1.22	
150.00	21.00	0.70	19.02	0.76	1.01	
180.00	20.00	0.67	18.11	0.72	0.96	
210.00	19.50	0.65	17.66	0.70	0.93	69.39
240.00	18.50	0.62	16.75	0.67	0.89	
300.00	36.50	0.61	16.53	0.66	0.87	
360.00	35.00	0.58	15.85	0.63	0.84	65.17
420.00	35.00	0.58	15.85	0.63	0.84	
480.00	32.00	0.53	14.49	0.58	0.77	67.98
540.00	29.50	0.49	13.36	0.53	0.71	
600.00	26.50	0.44	12.00	0.48	0.63	70.79
660.00	25.50	0.43	11.55	0.46	0.61	
720.00	23.50	0.39	10.64	0.42	0.56	70.79
1380.00	245.00	0.37	10.09	0.40	0.53	
1440.00	19.50	0.33	8.83	0.35	0.47	69.39
1500.00	19.00	0.32	8.60	0.34	0.45	

RUN NO. V
MEMBRANE: PVDF
PRESSURE: 130.00 psi
PRESSURE CORRECTION FACTOR 1.13

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	133.50					
90.00	40.50	1.35	36.68	1.46	1.65	69.39
120.00	30.00	1.00	27.17	1.08	1.22	
150.00	24.50	0.82	22.19	0.88	1.00	
180.00	23.00	0.77	20.83	0.83	0.94	
210.00	22.00	0.73	29.89	0.79	0.89	66.58
240.00	21.00	0.70	24.91	0.76	0.85	
300.00	38.50	0.64	24.45	0.69	0.78	
360.00	37.00	0.62	22.64	0.67	0.75	72.20
420.00	35.00	0.58	21.74	0.63	0.71	
480.00	32.50	0.54	20.83	0.59	0.66	67.98
540.00	30.00	0.50	19.47	0.54	0.61	
600.00	27.00	0.45	18.57	0.49	0.55	69.39
660.00	26.00	0.43	11.77	0.47	0.53	
720.00	24.00	0.40	10.87	0.43	0.49	69.39
1380.00	235.00	0.36	9.67	0.38	0.43	
1440.00	19.50	0.33	8.83	0.35	0.40	70.79
1500.00	19.00	0.32	8.60	0.34	0.39	

RUN NO. V
MEMBRANE: PS
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	40.00	1.33	36.23	1.44	1.92	70.79
120.00	30.00	1.00	27.17	1.08	1.44	
150.00	24.50	0.82	22.19	0.88	1.17	
180.00	23.00	0.77	20.83	0.83	1.10	
210.00	22.00	0.73	19.92	0.79	1.05	70.79
240.00	21.00	0.70	19.02	0.76	1.01	
300.00	39.00	0.65	17.66	0.70	0.93	
360.00	39.50	0.66	17.89	0.71	0.95	66.58
420.00	36.50	0.61	16.53	0.66	0.87	
480.00	34.00	0.57	15.40	0.61	0.81	67.98
540.00	31.50	0.53	14.26	0.57	0.75	
600.00	28.00	0.47	12.68	0.50	0.67	69.39
660.00	27.00	0.45	12.23	0.49	0.65	
720.00	26.00	0.43	11.77	0.47	0.62	67.98
1380.00	250.00	0.38	10.29	0.41	0.54	
1440.00	21.50	0.36	9.74	0.39	0.51	67.98
1500.00	21.00	0.35	9.51	0.38	0.50	

RUN NO. V
MEMBRANE: CA
PRESSURE: 140.00 psi
PRESSURE CORRECTION FACTOR 1.07

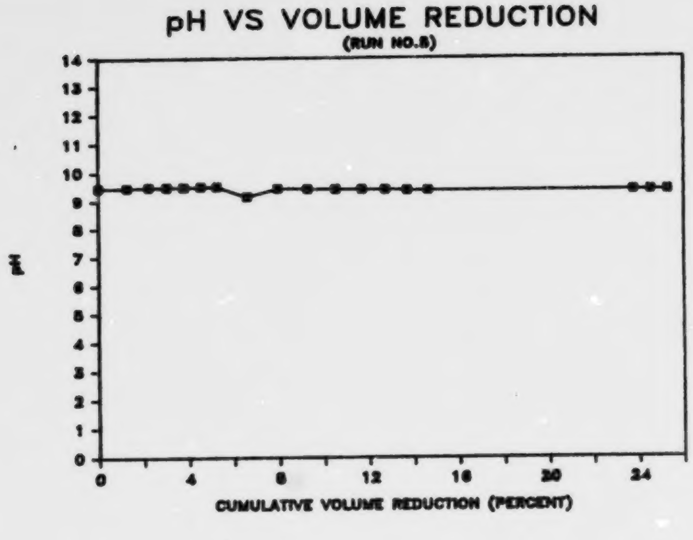
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	149.00					
90.00	59.50	1.98	53.89	2.14	2.29	69.39
120.00	45.50	1.52	41.21	1.64	1.75	
150.00	37.50	1.25	33.96	1.35	1.44	
180.00	35.00	1.17	31.70	1.26	1.35	
210.00	34.00	1.13	30.79	1.22	1.31	70.79
240.00	32.00	1.07	28.98	1.15	1.23	
300.00	61.50	1.03	27.85	1.11	1.18	
360.00	62.00	1.03	28.08	1.12	1.19	53.93
420.00	61.50	1.03	27.85	1.11	1.18	
480.00	58.00	0.97	26.26	1.04	1.12	67.98
540.00	56.00	0.93	25.36	1.01	1.08	
600.00	47.50	0.79	21.51	0.86	0.91	70.79
660.00	45.50	0.76	20.60	0.82	0.88	
720.00	42.50	0.71	19.25	0.77	0.82	69.39
1380.00	410.00	0.62	16.88	0.67	0.72	
1440.00	34.50	0.58	15.62	0.62	0.66	70.79
1500.00	34.00	0.57	15.40	0.61	0.65	

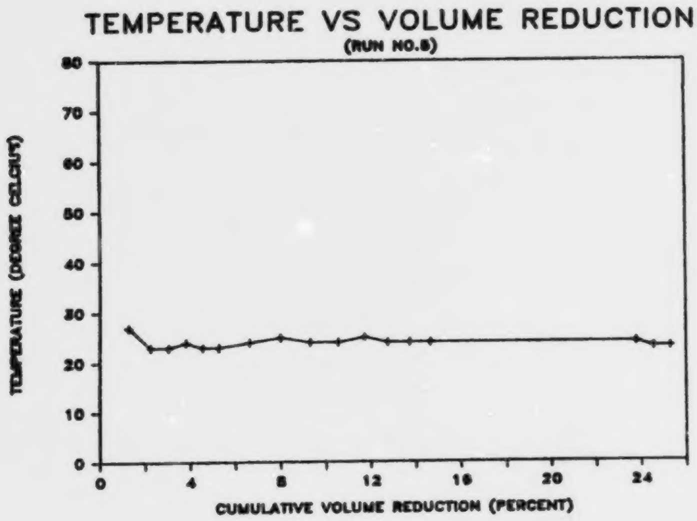
RUN NO. V
MEMBRANE: TFC
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

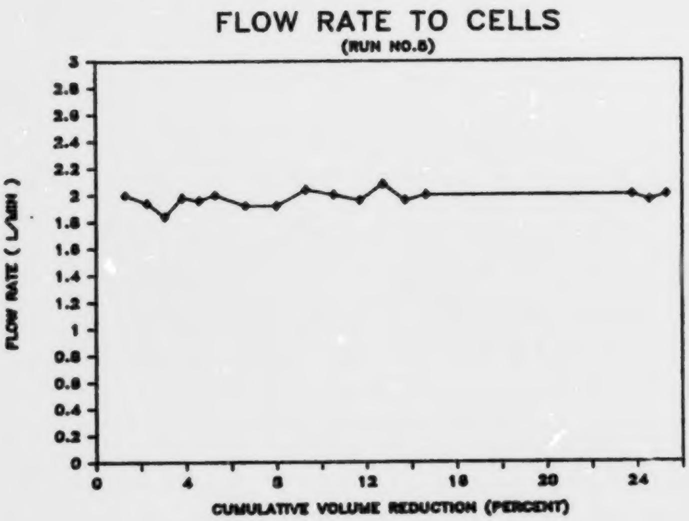
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	59.00					
90.00	36.00	1.20	32.60	1.30	1.30	72.20
120.00	28.00	0.93	25.36	1.01	1.01	
150.00	23.50	0.78	21.28	0.85	0.85	
180.00	23.50	0.78	21.28	0.85	0.85	
210.00	23.50	0.78	21.28	0.85	0.85	65.17
240.00	23.00	0.77	20.83	0.83	0.83	
300.00	45.00	0.75	20.38	0.81	0.81	
360.00	47.00	0.78	21.28	0.85	0.85	69.39
420.00	45.00	0.75	20.38	0.81	0.81	
480.00	42.00	0.70	19.02	0.76	0.76	67.98
540.00	39.50	0.66	17.89	0.71	0.71	
600.00	34.00	0.57	15.40	0.61	0.61	70.79
660.00	33.00	0.55	14.94	0.59	0.59	
720.00	30.00	0.50	13.59	0.54	0.54	67.98
1380.00	295.00	0.45	12.14	0.48	0.48	
1440.00	25.00	0.42	11.32	0.45	0.45	72.20
1500.00	24.50	0.41	11.09	0.44	0.44	

RUN NO. V
MEMBRANE: CELLULOSE
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORRECTED FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	17.00					
90.00	8.00	0.27	7.25	0.29	0.29	70.79
120.00	6.50	0.22	5.89	0.23	0.23	
150.00	6.00	0.20	5.43	0.22	0.22	
180.00	6.00	0.20	5.43	0.22	0.22	
210.00	6.00	0.20	5.43	0.22	0.22	67.98
240.00	6.00	0.20	5.43	0.22	0.22	
300.00	11.50	0.19	5.21	0.21	0.21	
360.00	13.00	0.22	5.89	0.23	0.23	66.58
420.00	13.50	0.23	6.11	0.24	0.24	
480.00	13.50	0.23	6.11	0.24	0.24	69.39
540.00	13.50	0.23	6.11	0.24	0.24	
600.00	12.00	0.20	5.43	0.22	0.22	67.98
660.00	12.00	0.20	5.43	0.22	0.22	
720.00	11.50	0.19	5.21	0.21	0.21	66.58
1380.00	127.00	0.19	5.23	0.21	0.21	
1440.00	11.50	0.19	5.21	0.21	0.21	67.98
1500.00	11.50	0.19	5.21	0.21	0.21	


pH VS VOLUME REDUCTION
(RUN NO.5)


TEMPERATURE VS VOLUME REDUCTION
(RUN NO.5)


FLOW RATE TO CELLS
(RUN NO.5)

FLUX RATES FOR EMULSION SCREENING TEST NO. 6

Temp. = 45°C pH = 9.5 Pressure = 100 - 150 psi

TIME (MIN)	CUMULATIVE VOLUME REDUCTION (%)	CHP FLUX RATE M3/M2·D	PS FLUX RATE M3/M2·D	PVDF FLUX RATE M3/M2·D	CA FLUX RATE M3/M2·D	TFC FLUX RATE M3/M2·D	CELLULOSE FLUX RATE M3/M2·D
60.00	0.00						
90.00	1.54	2.15	2.25	1.95	2.43	1.67	0.47
120.00	2.75	1.65	1.72	1.51	1.89	1.44	0.41
150.00	4.03	1.70	1.77	1.53	2.00	1.58	0.45
180.00	5.08	1.36	1.46	1.24	1.64	1.30	0.40
210.00	6.08	1.29	1.39	1.18	1.56	1.26	0.38
240.00	7.04	1.17	1.32	1.12	1.54	1.26	0.40
300.00	8.94	1.23	1.32	1.08	1.41	1.24	0.41
360.00	10.59	1.08	1.14	0.91	1.30	1.06	0.35
420.00	12.17	1.04	1.09	0.81	1.22	1.04	0.35
480.00	13.76	1.03	1.09	0.92	1.20	1.02	0.35
540.00	15.25	0.98	1.03	0.80	1.13	0.95	0.34
600.00	16.75	1.03	1.04	0.81	1.10	0.97	0.36
660.00	18.23	1.01	1.02	0.79	1.10	0.93	0.36
720.00	19.69	1.01	1.01	0.78	1.09	0.91	0.38
1440.00	35.99	0.92	0.92	0.70	1.00	0.85	0.39
1500.00	37.27	0.80	0.85	0.65	0.95	0.81	0.41

ANALYTICAL DATA FOR SCREENING TEST NO.6

PARAMETER	UNITS	INIT FEED	FINAL FEED	MEAN FEED	C PERM	C %REJ	CA PERM	CA %REJ	PS PERM	PS %REJ	CHP PERM	CHP %REJ	TFC PERM	TFC %REJ	PVDF PERM	PVDF %REJ
pH		9.72	9.27	9.50	9.10		8.98		8.88		8.91		8.67		8.81	
COND	uHmos/cm	1050	1075	1063	1075	-1.2	800	24.7	590	44.5	610	42.6	290	72.7	545	48.7
TURB	NTU	980	2100	1540	1.15	99.9	0.92	99.9	0.63	100.0	0.85	99.9	0.22	100.0	0.58	100.0
TOC	mg/L	1400	1525	1463	139.5	90.5	32.0	97.8	58.5	96.0	73.0	95.0	35.5	97.6	52.0	96.4
TDS	mg/L	1040	NA	1040	845	18.8	515	50.5	460	55.8	440	57.7	245	76.4	415	60.1
HARDNESS	mg/L	8.47	NA	8.47	3.65	56.9	<0.15	100.0	<0.15	100.0	<0.15	100.0	<0.15	100.0	<0.15	100.0
DIS SIO2	mg/L	80.3	NA	80.3	58.4	27.3	60.8	24.3	47.0	41.5	58.4	27.3	49.7	38.1	58.2	27.5
TSS	mg/L	1610	NA	1610	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA

RUN NO.6
% CUMULATIVE VOLUME REDUCTION
SEPTEMBER 6 1984 VOLUME OF FEED 17100.00 MLS

TIME (MIN)	CHP PERM VOL (MLS)	PS PERM VOL (MLS)	PVDF PERM VOL (MLS)	CA PERM VOL (MLS)	TFC PERM VOL (MLS)	CELLULOSE PERM VOL (MLS)	TOTAL PERM VOL (MLS)	CONC pH	CONC TEMP (C)	CUM VOL RED (%)	FLOW TO CELLS (L/MIN)
15.00	28.50	46.00	47.50	44.50	12.50	3.50	0.00		30.00	0.00	2.00
30.00	30.50	34.50	35.00	43.50	21.50	6.50	0.00		40.00	0.00	2.08
45.00	29.00	30.50	31.50	41.00	24.50	6.50	0.00		44.00	0.00	1.80
60.00	21.00	24.50	25.00	32.00	20.50	6.00	0.00	9.57	43.00	0.00	1.88
90.00	45.00	47.00	48.00	63.00	46.50	13.00	262.50	9.55	50.00	1.54	2.08
120.00	34.50	36.00	37.00	49.00	40.00	11.50	208.00	9.40	47.00	2.75	2.04
150.00	35.50	37.00	37.50	52.00	44.00	12.50	218.50	9.39	46.00	4.03	2.04
180.00	28.50	30.50	30.50	42.50	36.00	11.00	179.00	9.39	43.00	5.08	2.00
210.00	27.00	29.00	29.00	40.50	35.00	10.50	171.00	9.37	44.00	6.08	1.94
240.00	24.50	27.50	27.50	40.00	35.00	11.00	165.50	9.36	45.00	7.04	2.12
300.00	51.50	55.00	53.00	73.00	69.00	22.50	324.00	9.38	46.00	8.94	2.16
360.00	45.00	47.50	44.50	67.50	59.00	19.50	283.00	9.35	42.00	10.59	2.00
420.00	43.50	45.50	40.00	63.50	58.00	19.50	270.00	9.34	46.00	12.17	2.00
480.00	43.00	45.50	45.00	62.50	56.60	19.50	272.10	9.33	45.00	13.76	1.96
540.00	41.00	43.00	39.50	58.80	52.50	19.00	253.80	9.34	44.00	15.25	1.84
600.00	43.00	43.50	40.00	57.00	54.00	20.00	257.50	9.34	46.00	16.75	1.84
660.00	42.00	42.50	39.00	57.00	51.50	20.00	252.00	9.33	44.00	18.23	1.80
720.00	42.00	42.00	38.50	56.50	50.50	21.00	250.50	9.34	46.00	19.69	1.88
1440.00	460.00	460.00	414.00	624.00	566.00	263.00	2787.00	9.26	47.00	35.99	1.88
1500.00	33.50	35.50	32.00	49.50	45.00	23.00	218.50	9.27	49.00	37.27	2.20

RUN NO. VI
MEMBRANE: CHP
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	45.00	1.50	40.76	1.62	2.15	70.79
120.00	34.50	1.15	31.25	1.24	1.65	
150.00	35.50	1.18	32.15	1.28	1.70	
180.00	28.50	0.95	25.81	1.03	1.36	
210.00	27.00	0.90	24.45	0.97	1.29	69.39
240.00	24.50	0.82	22.19	0.88	1.17	
300.00	51.50	0.86	23.32	0.93	1.23	
360.00	45.00	0.75	20.38	0.81	1.08	69.39
420.00	43.50	0.73	19.70	0.78	1.04	
480.00	43.00	0.72	19.47	0.77	1.03	69.39
540.00	41.00	0.68	18.57	0.74	0.98	
600.00	43.00	0.72	19.47	0.77	1.03	75.01
660.00	42.00	0.70	19.02	0.76	1.01	
720.00	42.00	0.70	19.02	0.76	1.01	72.20
1440.00	460.00	0.64	17.36	0.69	0.92	
1500.00	33.50	0.56	15.17	0.60	0.80	72.20

RUN NO. VI
MEMBRANE: PVDF
PRESSURE: 130.00 psi
PRESSURE CORRECTION FACTOR 1.13

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	133.50					
90.00	48.00	1.60	43.47	1.73	1.95	70.79
120.00	37.00	1.23	33.51	1.33	1.51	
150.00	37.50	1.25	33.96	1.35	1.53	
180.00	30.50	1.02	27.62	1.10	1.24	
210.00	29.00	0.97	29.89	1.04	1.18	73.60
240.00	27.50	0.92	24.91	0.99	1.12	
300.00	53.00	0.88	24.45	0.95	1.08	
360.00	44.50	0.74	22.64	0.80	0.91	72.20
420.00	40.00	0.67	21.74	0.72	0.81	
480.00	45.00	0.75	20.83	0.81	0.92	67.98
540.00	39.50	0.66	19.47	0.71	0.80	
600.00	40.00	0.67	18.57	0.72	0.81	72.20
660.00	39.00	0.65	17.66	0.70	0.79	
720.00	38.50	0.64	17.43	0.69	0.78	69.39
1440.00	414.00	0.58	15.62	0.62	0.70	
1500.00	32.00	0.53	14.49	0.58	0.65	75.01

RUN NO. VI
MEMBRANE: PS
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	47.00	1.57	42.57	1.69	2.25	72.20
120.00	36.00	1.20	32.60	1.30	1.72	
150.00	37.00	1.23	33.51	1.33	1.77	
180.00	30.50	1.02	27.62	1.10	1.46	
210.00	29.00	0.97	26.26	1.04	1.39	70.79
240.00	27.50	0.92	24.91	0.99	1.32	
300.00	55.00	0.92	24.91	0.99	1.32	
360.00	47.50	0.79	21.51	0.86	1.14	72.20
420.00	45.50	0.76	20.60	0.82	1.09	
480.00	45.50	0.76	20.60	0.82	1.09	72.20
540.00	43.00	0.72	19.47	0.77	1.03	
600.00	43.50	0.73	19.70	0.78	1.04	73.60
660.00	42.50	0.71	19.25	0.77	1.02	
720.00	42.00	0.70	19.02	0.76	1.01	70.79
1440.00	460.00	0.64	17.36	0.69	0.92	
1500.00	35.50	0.59	16.08	0.64	0.85	72.20

RUN NO. V
MEMBRANE: CA
PRESSURE: 140.00 psi
PRESSURE CORRECTION FACTOR 1.07

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	149.00					
90.00	63.00	2.10	57.06	2.27	2.43	70.79
120.00	49.00	1.63	44.38	1.76	1.89	
150.00	52.00	1.73	47.09	1.87	2.00	
180.00	42.50	1.42	38.49	1.53	1.64	
210.00	40.50	1.35	36.68	1.46	1.56	67.98
240.00	40.00	1.33	36.23	1.44	1.54	
300.00	73.00	1.22	33.06	1.31	1.41	
360.00	67.50	1.13	30.57	1.22	1.30	70.79
420.00	63.50	1.06	28.75	1.14	1.22	
480.00	62.50	1.04	28.30	1.13	1.20	70.79
540.00	58.80	0.98	26.63	1.06	1.13	
600.00	57.00	0.95	25.81	1.03	1.10	73.60
660.00	57.00	0.95	25.81	1.03	1.10	
720.00	56.50	0.94	25.59	1.02	1.09	72.20
1440.00	624.00	0.87	23.55	0.94	1.00	
1500.00	49.50	0.83	22.42	0.89	0.95	70.79

RUN NO. VI
MEMBRANE: TFC
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

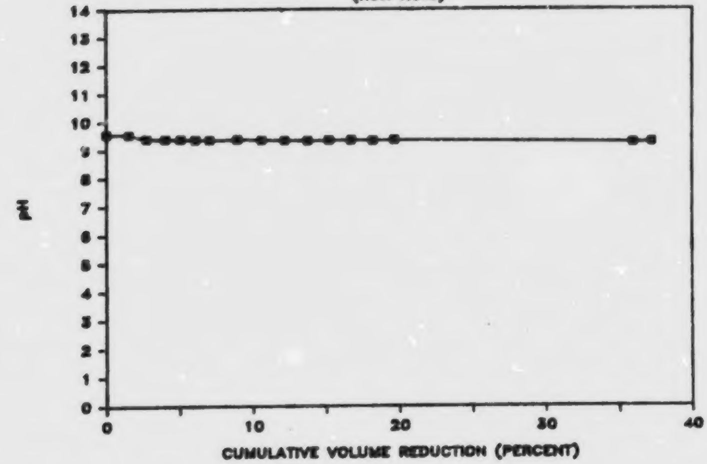
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	59.00					
90.00	46.50	1.55	42.11	1.67	1.67	70.79
120.00	40.00	1.33	36.23	1.44	1.44	
150.00	44.00	1.47	39.85	1.58	1.58	
180.00	36.00	1.20	32.60	1.30	1.30	
210.00	35.00	1.17	31.70	1.26	1.26	72.20
240.00	35.00	1.17	31.70	1.26	1.26	
300.00	69.00	1.15	31.25	1.24	1.24	
360.00	59.00	0.98	26.72	1.06	1.06	72.20
420.00	58.00	0.97	26.26	1.04	1.04	
480.00	56.60	0.94	25.63	1.02	1.02	75.01
540.00	52.50	0.88	23.77	0.95	0.95	
600.00	54.00	0.90	24.45	0.97	0.97	72.20
660.00	51.50	0.86	23.32	0.93	0.93	
720.00	50.50	0.84	22.87	0.91	0.91	73.60
1440.00	566.00	0.79	21.36	0.85	0.85	
1500.00	45.00	0.75	20.38	0.81	0.81	73.60

RUN NO. VI
MEMBRANE: CELLULOSE
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	17.00					
90.00	13.00	0.43	11.77	0.47	0.47	69.39
120.00	11.50	0.38	10.42	0.41	0.41	
150.00	12.50	0.42	11.32	0.45	0.45	
180.00	11.00	0.37	9.96	0.40	0.40	
210.00	10.50	0.35	9.51	0.38	0.38	70.79
240.00	11.00	0.37	9.96	0.40	0.40	
300.00	22.50	0.38	10.19	0.41	0.41	
360.00	19.50	0.33	8.83	0.35	0.35	70.79
420.00	19.50	0.33	8.83	0.35	0.35	
480.00	19.50	0.33	8.83	0.35	0.35	73.60
540.00	19.00	0.32	8.60	0.34	0.34	
600.00	20.00	0.33	9.06	0.36	0.36	70.79
660.00	20.00	0.33	9.06	0.36	0.36	
720.00	21.00	0.35	9.51	0.38	0.38	69.39
1440.00	263.00	0.37	9.92	0.39	0.39	
1500.00	23.00	0.38	10.42	0.41	0.41	70.79

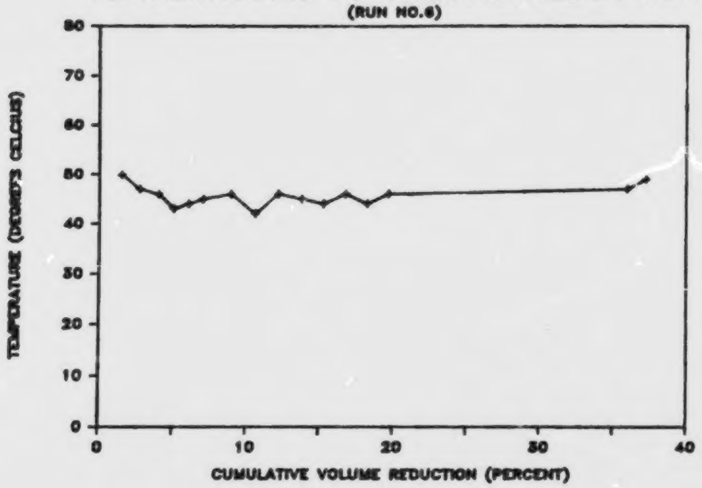
pH VS VOLUME REDUCTION

(RUN NO.6)



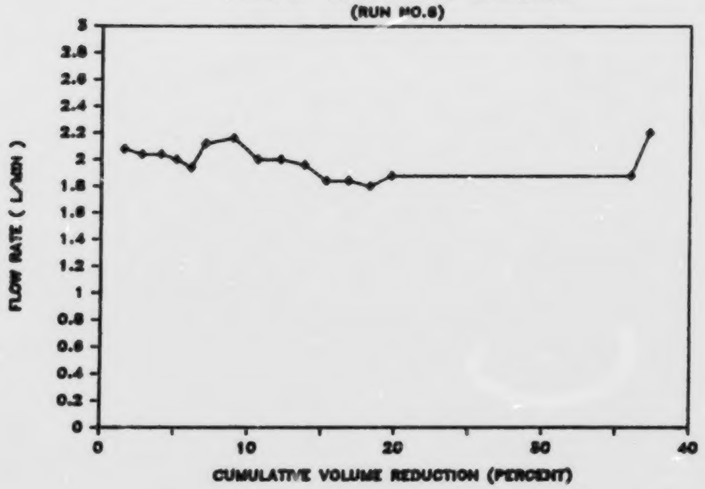
TEMPERATURE VS VOLUME REDUCTION

(RUN NO.6)



FLOW RATE TO CELLS

(RUN NO.6)



FLUX RATES FOR EMULSION SCREENING TEST NO. 7

Temp. = 65°C pH = as is (7.5 - 8.0) Pressure = 100 - 150 psi

TIME (MIN)	CUMULATIVE VOLUME REDUCTION (%)	CHP FLUX RATE M3/M2·D	PS FLUX RATE M3/M2·D	PVDF FLUX RATE M3/M2·D	CA FLUX RATE M3/M2·D	TFC FLUX RATE M3/M2·D	CHP2 FLUX RATE M3/M2·D
60.00	0.00						
90.00	1.88	1.44	2.49	2.16	2.81	2.20	1.89
120.00	3.45	1.36	2.01	1.73	2.33	1.93	1.51
150.00	4.80	1.10	1.75	1.51	1.93	1.69	1.31
180.00	5.97	0.96	1.53	1.32	1.64	1.48	1.15
210.00	7.12	0.96	1.53	1.30	1.58	1.48	1.13
240.00	8.44	1.15	1.75	1.48	1.75	1.69	1.30
300.00	10.92	1.15	1.68	1.39	1.58	1.59	1.22
360.00	12.64	0.79	1.17	0.97	1.11	1.08	0.84
420.00	14.62	0.96	1.38	1.11	1.24	1.24	0.95
480.00	16.37	0.78	1.23	1.00	1.12	1.11	0.86
540.00	17.84	0.63	0.95	0.79	0.94	0.95	0.77
600.00	19.47	0.74	1.09	0.89	1.03	1.05	0.85
660.00	21.12	0.79	1.13	0.89	1.02	1.04	0.85
720.00	22.73	0.78	1.11	0.88	0.98	1.02	0.83
1380.00	39.81	0.84	1.07	0.85	0.92	0.95	0.79
1440.00	41.30	0.85	0.96	0.83	0.89	0.90	0.77
1500.00	42.58	0.67	0.78	0.71	0.79	0.79	0.69

ANALYTICAL DATA FOR SCREENING TEST NO.7

PARAMETER	UNITS	INIT FEED	FINAL FEED	MEAN FEED	CHP PERM	CHP %REJ	CA PERM	CA %REJ	PS PERM	PS %REJ	CHP PERM	CHP %REJ	TFC PERM	TFC %REJ	PVDF PERM	PVDF %REJ
pH		8.15	9.48	8.82	8.63		8.93		8.81		8.94		8.9		8.85	
COND	uHmos/cm	975	1400	1188	560	52.8	630	46.9	580	51.2	620	47.8	375	68.4	590	50.3
TURB	NTU	520	4300	2410	0.54	100.0	0.43	100.0	0.41	100.0	0.40	100.0	0.29	100.0	0.5	100.0
TOC	mg/L	2050	3788	2919	58.5	98.0	69.5	97.6	47.0	98.4	59.5	98.0	28.0	99.0	53.0	98.2
TDS	mg/L	987.5	NA	987.5	382	61.3	561	43.2	398.5	59.6	469.5	52.5	261.5	73.5	406.5	58.8
HARDNESS	mg/L	21.67	NA	21.67	<0.15	100.0	<0.15	100.0	<0.15	100.0	<0.15	100.0	<0.15	100.0	<0.15	100.0
DIS SIO2	mg/L	78.6	NA	78.6	64.3	18.2	64.2	18.3	63.3	19.5	61.9	21.2	64.2	18.3	63.7	19.0
TSS	mg/L	1616	NA	1616	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA

RUN NO.7
% VOLUME REDUCTION
SEPTEMBER 17 1984 VOLUME OF FEED 17100.00 MLS

TIME (MIN)	CHP1 PERM VOL (MLS)	PS PERM VOL (MLS)	PVDF PERM VOL (MLS)	CA PERM VOL (MLS)	TFC PERM VOL (MLS)	CHP2 PERM VOL (MLS)	TOTAL PERM VOL (MLS)	FEED pH	FEED TEMP (C)	CUM VOL RED (%)	FLOW TO CELLS (L/MIN)
15.00	12.00	37.00	38.50	33.00	14.00	39.50	0.00			0.00	2.00
30.00	13.50	26.50	28.00	36.00	22.50	35.00	0.00			0.00	2.08
45.00	13.00	25.00	27.00	36.00	24.50	30.00	0.00			0.00	1.80
60.00	13.50	24.00	26.50	35.50	27.00	27.50	0.00		68.00	0.00	1.88
90.00	30.00	52.00	53.00	73.00	61.00	52.50	321.50	8.15	61.00	1.88	2.08
120.00	28.50	42.00	42.50	60.50	53.50	42.00	269.00		67.00	3.45	2.04
150.00	23.00	36.50	37.00	50.00	47.00	36.30	229.80		58.00	4.80	2.04
180.00	20.00	32.00	32.50	42.50	41.00	32.00	200.00		66.00	5.97	2.00
210.00	20.00	32.00	32.00	41.00	41.00	31.50	197.50		68.00	7.12	1.94
240.00	24.00	36.50	36.50	45.50	47.00	36.00	225.50		69.00	8.44	2.12
300.00	48.00	70.00	68.50	82.00	88.50	67.50	424.50		59.00	10.92	2.16
360.00	33.00	49.00	47.50	57.50	60.00	46.50	293.50		65.00	12.64	2.00
420.00	40.00	57.50	54.50	64.50	69.00	53.00	338.50		61.00	14.62	2.00
480.00	32.50	51.50	49.00	58.00	61.50	47.50	300.00		60.00	16.37	1.96
540.00	26.50	39.50	39.00	49.00	53.00	43.00	250.00		67.00	17.84	1.84
600.00	31.00	45.50	44.00	53.50	58.50	47.00	279.50		64.00	19.47	1.84
660.00	33.00	47.00	44.00	53.00	58.00	47.00	282.00		63.00	21.12	1.80
720.00	32.50	46.50	43.50	51.00	56.50	46.00	276.00			22.73	1.88
1380.00	385.00	490.00	457.00	528.00	578.00	482.00	2920.00			39.81	1.88
1440.00	35.50	40.00	41.00	46.00	50.00	42.50	255.00		60.00	41.30	2.20
1500.00	28.00	32.50	35.00	41.00	44.00	38.50	219.00	9.48	60.00	42.58	2.20

RUN NO. VII
MEMBRANE: CHP1
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00						
90.00	30.00	1.00	27.17	1.08	1.44	
120.00	28.50	0.95	25.81	1.03	1.36	67.39
150.00	23.00	0.77	20.83	0.83	1.10	
180.00	20.00	0.67	18.11	0.72	0.96	
210.00	20.00	0.67	18.11	0.72	0.96	
240.00	24.00	0.80	21.74	0.86	1.15	69.39
300.00	48.00	0.80	21.74	0.86	1.15	
360.00	33.00	0.55	14.94	0.59	0.79	70.79
420.00	40.00	0.67	18.11	0.72	0.96	
480.00	32.50	0.54	14.72	0.59	0.78	66.58
540.00	26.50	0.44	12.00	0.48	0.63	
600.00	31.00	0.52	14.04	0.56	0.74	69.39
660.00	33.00	0.55	14.94	0.59	0.79	
720.00	32.50	0.54	14.72	0.59	0.78	67.98
1380.00	385.00	0.58	15.85	0.63	0.84	
1440.00	35.50	0.59	16.08	0.64	0.85	65.17
1500.00	28.00	0.47	12.68	0.50	0.67	

RUN NO. VII
MEMBRANE: PVDF
PRESSURE: 130.00 psi
PRESSURE CORRECTION FACTOR 1.13

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	133.50					
90.00	53.00	1.77	48.00	1.91	2.16	
120.00	42.50	1.42	38.49	1.53	1.73	65.17
150.00	37.00	1.23	33.51	1.33	1.51	
180.00	32.50	1.08	29.43	1.17	1.32	
210.00	32.00	1.07	29.89	1.15	1.30	
240.00	36.50	1.22	24.91	1.31	1.48	67.98
300.00	68.50	1.14	24.45	1.23	1.39	
360.00	47.50	0.79	22.64	0.86	0.97	73.60
420.00	54.50	0.91	21.74	0.98	1.11	
480.00	49.00	0.82	20.83	0.88	1.00	70.79
540.00	39.00	0.65	19.47	0.70	0.79	
600.00	44.00	0.73	18.57	0.79	0.89	63.77
660.00	44.00	0.73	19.92	0.79	0.89	
720.00	43.50	0.73	19.70	0.78	0.88	67.98
1380.00	457.00	0.69	18.81	0.75	0.85	
1440.00	41.00	0.68	18.57	0.74	0.83	67.98
1500.00	35.00	0.58	15.85	0.63	0.71	

RUN NO. VII
MEMBRANE: PS
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	118.50					
90.00	52.00	1.73	47.09	1.87	2.49	
120.00	42.00	1.40	38.04	1.51	2.01	63.77
150.00	36.50	1.22	33.06	1.31	1.75	
180.00	32.00	1.07	28.98	1.15	1.53	
210.00	32.00	1.07	28.98	1.15	1.53	
240.00	36.50	1.22	33.06	1.31	1.75	67.98
300.00	70.00	1.17	31.70	1.26	1.68	
360.00	49.00	0.82	22.19	0.88	1.17	73.60
420.00	57.50	0.96	26.04	1.04	1.38	
480.00	51.50	0.86	23.32	0.93	1.23	67.98
540.00	39.50	0.66	17.89	0.71	0.95	
600.00	45.50	0.76	20.60	0.82	1.09	63.77
660.00	47.00	0.78	21.28	0.85	1.13	
720.00	46.50	0.78	21.06	0.84	1.11	67.98
1380.00	490.00	0.74	20.17	0.80	1.07	
1440.00	40.00	0.67	18.11	0.72	0.96	67.98
1500.00	32.50	0.54	14.72	0.59	0.78	

RUN NO. VII
MEMBRANE: CA
PRESSURE: 140.00 psi
PRESSURE CORRECTION FACTOR 1.07

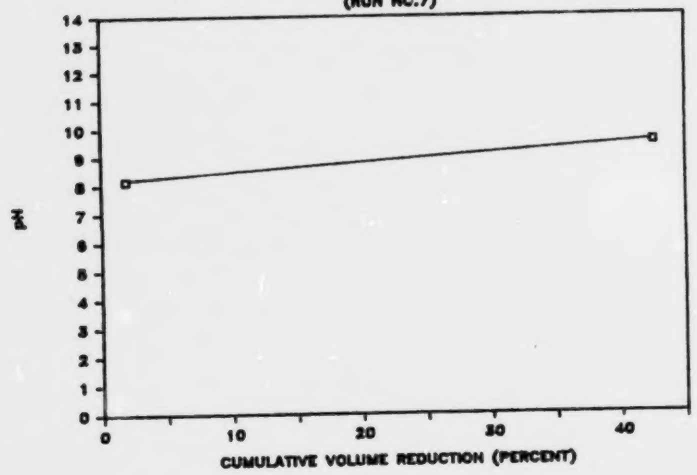
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	149.00					
90.00	73.00	2.43	66.11	2.63	2.81	
120.00	60.50	2.02	54.79	2.18	2.33	67.98
150.00	50.00	1.67	45.28	1.80	1.93	
180.00	42.50	1.42	38.49	1.53	1.64	
210.00	41.00	1.37	37.13	1.48	1.58	
240.00	45.50	1.52	41.21	1.64	1.75	69.39
300.00	82.00	1.37	37.13	1.48	1.58	
360.00	57.50	0.96	26.04	1.04	1.11	70.79
420.00	64.50	1.08	29.21	1.16	1.24	
480.00	58.00	0.97	26.26	1.04	1.12	67.98
540.00	49.00	0.82	22.19	0.88	0.94	
600.00	53.50	0.89	24.23	0.96	1.03	67.98
660.00	53.00	0.88	24.00	0.95	1.02	
720.00	51.00	0.85	23.09	0.92	0.98	67.98
1380.00	528.00	0.80	21.74	0.86	0.92	
1440.00	46.00	0.77	20.83	0.83	0.89	67.98
1500.00	41.00	0.68	18.57	0.74	0.79	

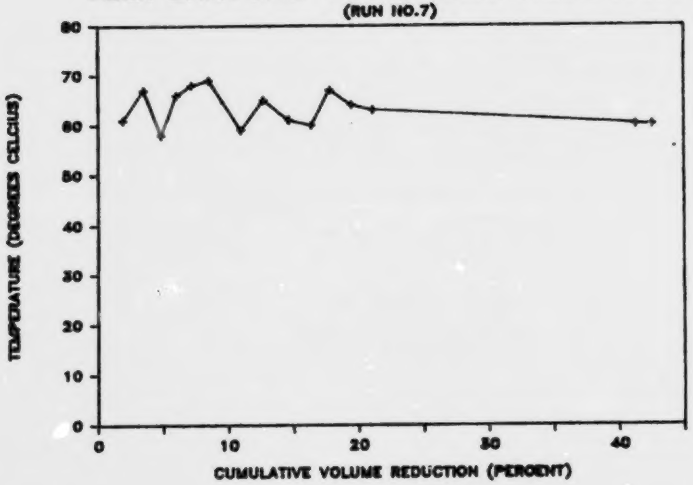
RUN NO. VII
MEMBRANE: TFC
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

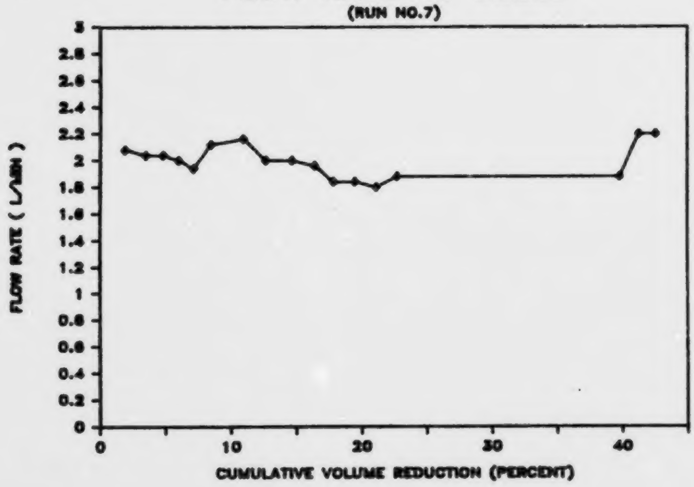
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	59.00					
90.00	61.00	2.03	55.25	2.20	2.20	
120.00	53.50	1.78	48.45	1.93	1.93	63.77
150.00	47.00	1.57	42.57	1.69	1.69	
180.00	41.00	1.37	37.13	1.48	1.48	
210.00	41.00	1.37	37.13	1.48	1.48	
240.00	47.00	1.57	42.57	1.69	1.69	53.93
300.00	88.50	1.48	40.08	1.59	1.59	
360.00	60.00	1.00	27.17	1.08	1.08	70.79
420.00	69.00	1.15	31.25	1.24	1.24	
480.00	61.50	1.03	27.85	1.11	1.11	72.20
540.00	53.00	0.88	24.00	0.95	0.95	
600.00	58.50	0.98	26.49	1.05	1.05	70.79
660.00	58.00	0.97	26.26	1.04	1.04	
720.00	56.50	0.94	25.59	1.02	1.02	70.79
1380.00	578.00	0.88	23.79	0.95	0.95	
1440.00	50.00	0.83	22.64	0.90	0.90	63.77
1500.00	44.00	0.73	19.92	0.79	0.79	

RUN NO. VII
MEMBRANE: CHP2
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	SURFACE TENSION (DYNES/CM)
60.00	17.00					
90.00	52.50	1.75	47.55	1.89	1.89	
120.00	42.00	1.40	38.04	1.51	1.51	69.39
150.00	36.30	1.21	32.88	1.31	1.31	
180.00	32.00	1.07	28.98	1.15	1.15	
210.00	31.50	1.05	28.53	1.13	1.13	
240.00	36.00	1.20	32.60	1.30	1.30	69.39
300.00	67.50	1.13	30.57	1.22	1.22	
360.00	46.50	0.78	21.06	0.84	0.84	73.60
420.00	53.00	0.88	24.00	0.95	0.95	
480.00	47.50	0.79	21.51	0.86	0.86	67.98
540.00	43.00	0.72	19.47	0.77	0.77	
600.00	47.00	0.78	21.28	0.85	0.85	59.55
660.00	47.00	0.78	21.28	0.85	0.85	
720.00	46.00	0.77	20.83	0.83	0.83	69.39
1380.00	482.00	0.73	19.84	0.79	0.79	
1440.00	42.50	0.71	19.25	0.77	0.77	63.77
1500.00	38.50	0.64	17.43	0.69	0.69	


pH VS VOLUME REDUCTION
(RUN NO.7)


TEMPERATURE VS VOLUME REDUCTION
(RUN NO.7)


FLOW RATE TO CELLS
(RUN NO.7)

FLUX RATES FOR EMULSION SCREENING TEST NO. 8

Temp. = 65°C pH = as is (7.5 - 8.0) Pressure = 100 - 150 psi

TIME (MIN)	CUMULATIVE VOLUME REDUCTION (%)	CELL1 PS FLUX RATE M3/M2·D	CELL2 PS FLUX RATE M3/M2·D	CELL3 PS FLUX RATE M3/M2·D	CELL4 PS FLUX RATE M3/M2·D	CELL5 PS FLUX RATE M3/M2·D	CELL6 PS FLUX RATE M3/M2·D
60.00	1.58	2.51	2.84	1.84	2.03	1.71	1.82
120.00	2.78	1.72	1.80	1.42	1.51	1.47	1.46
180.00	3.87	1.53	1.60	1.30	1.34	1.38	1.33
240.00	4.89	1.51	1.51	1.15	1.26	1.27	1.25
1260.00	11.78	0.54	0.58	0.46	0.49	0.54	0.52
1380.00	12.94	0.87	0.89	0.64	0.71	0.74	0.70
1500.00	14.05	0.80	0.84	0.61	0.68	0.70	0.68
1620.00	15.15	0.83	0.86	0.59	0.69	0.65	0.68
2640.00	21.78	0.58	0.63	0.38	0.51	0.43	0.52
2760.00	22.45	0.57	0.60	0.30	0.45	0.32	0.44
2880.00	23.11	0.56	0.60	0.28	0.41	0.31	0.45
3000.00	23.78	0.57	0.62	0.28	0.44	0.30	0.42
3120.00	24.33	0.49	0.52	0.23	0.38	0.24	0.35
4500.00	30.45	0.49	0.27	0.39	0.36	0.22	0.33
4538.00	30.61	0.49	0.25	0.40	0.32	0.23	0.31
5520.00	35.23	0.58	0.62	0.17	0.41	0.23	0.28
5640.00	35.76	0.57	0.61	0.16	0.39	0.22	0.24
5760.00	36.25	0.53	0.56	0.14	0.36	0.20	0.21
5940.00	36.86	0.44	0.47	0.11	0.30	0.16	0.17
6000.00	37.07	0.48	0.50	0.13	0.22	0.18	0.20
7405.00	40.92	0.37	0.11	0.29	0.15	0.12	0.22
7435.00	40.99	0.29	0.10	0.25	0.12	0.10	0.18
8760.00	44.54	0.37	0.39	0.08	0.25	0.12	0.13
8820.00	44.72	0.42	0.43	0.08	0.26	0.13	0.14
9900.00	47.95	0.42	0.45	0.08	0.27	0.13	0.13
10020.00	48.35	0.50	0.53	0.11	0.32	0.07	0.16

ANALYTICAL DATA FOR SCREENING TEST NO.8

PARAMETER	UNITS	INIT FEED	FINAL FEED	MEAN FEED	PS COM PERM	PS %REJ
pH		7.4	9.15	8.28	7.88	
COND	uHmos/cm	1000	1450	1225	580	52.7
TURB	NTU	1000	2350	1675	2.30	99.9
TOC	mg/L	158	240	199	46.0	76.9
TDS	mg/L	1154	492	1154	NA	100.0
HARDNESS	mg/L	4.75	NA	4.75	<0.15	100.0
DIS SIO2	mg/L	NA	NA	NA	NA	NA
TSS	mg/L	NA	NA	NA	NA	NA

RUN NO.8
% CUMULATIVE VOLUME REDUCTION
SEPTEMBER 17 1984 VOLUME OF FEED 38000.00 MLS

TIME (MIN)	CELL 1 PERM VOL (MLS)	CELL 2 PERM VOL (MLS)	CELL 3 PERM VOL (MLS)	CELL 4 PERM VOL (MLS)	CELL 5 PERM VOL (MLS)	CELL6 PERM VOL (MLS)	TOTAL PERM VOL (MLS)	FEED pH	FEED TEMP (C)	CUM VOL RED (%)	FLOW TO CELLS (L/MIN)
0.00											
60.00	105.00	103.00	100.00	89.00	102.00	101.00	600.00		24.00	1.58	2.16
120.00	72.00	75.00	74.00	76.50	79.00	81.00	457.50		39.00	2.78	2.00
180.00	64.00	67.00	66.00	71.50	72.00	74.00	414.50			3.87	2.12
240.00	63.00	63.00	62.00	66.00	64.00	69.50	387.50			4.89	2.08
1260.00	385.00	412.00	410.00	477.00	438.00	493.00	2615.00		39.00	11.78	2.04
1380.00	72.50	74.50	70.00	77.00	71.00	78.00	443.00		54.00	12.94	1.96
1500.00	66.50	70.50	66.50	72.50	68.00	76.00	420.00		65.00	14.05	2.04
1620.00	69.50	72.00	68.00	68.00	65.00	76.00	418.50		65.00	15.15	2.00
2640.00	414.00	450.00	428.00	376.00	359.00	492.00	2519.00		51.00	21.78	2.16
2760.00	47.50	50.00	44.00	33.00	33.00	48.50	256.00		63.00	22.45	1.80
2880.00	47.00	50.50	40.50	32.00	31.50	50.00	251.50		63.00	23.11	2.00
3000.00	48.00	52.00	43.00	31.00	31.00	47.00	252.00		66.00	23.78	1.80
3120.00	41.00	43.50	37.00	24.50	25.00	39.00	210.00		57.00	24.33	1.84
4500.00	468.00	256.00	412.00	262.00	500.00	427.00	2325.00		64.00	30.45	2.08
4538.00	13.00	6.50	10.00	7.50	14.00	11.00	62.00		68.00	30.61	2.00
5520.00	398.00	421.00	333.00	193.00	157.00	252.00	1754.00		60.00	35.23	2.00
5640.00	48.00	51.00	38.50	23.00	17.50	26.50	204.50		56.00	35.76	2.06
5760.00	44.00	46.50	35.00	20.50	15.50	23.50	185.00		67.00	36.25	1.94
5940.00	55.50	59.50	44.00	25.50	19.00	29.00	232.50		65.00	36.86	1.92
6000.00	20.00	21.00	11.00	9.50	7.00	11.00	79.50		69.00	37.07	2.00
7405.00	364.00	108.00	174.00	146.00	383.00	288.00	1463.00		59.00	40.92	1.80
7435.00	6.00	2.00	3.00	2.50	7.00	5.00	25.50		59.00	40.99	2.00
8760.00	340.00	362.00	266.00	133.00	94.00	154.00	1349.00		60.00	44.54	2.00
8820.00	17.50	18.00	13.00	7.00	4.60	7.50	67.60		52.00	44.72	2.00
9900.00	319.00	338.00	241.00	117.00	82.00	131.00	1228.00		60.00	47.95	2.00
10020.00	41.50	44.50	31.50	7.00	12.00	18.00	154.50		60.00	48.35	2.00

RUN NO. VIII
MEMBRANE: PS1
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)
60.00	105.00	1.75	47.55	1.89	2.51
120.00	72.00	1.20	32.60	1.30	1.72
180.00	64.00	1.07	28.98	1.15	1.53
240.00	63.00	1.05	28.53	1.13	1.51
1260.00	385.00	0.38	10.26	0.41	0.54
1380.00	72.50	0.60	16.42	0.65	0.87
1500.00	66.50	0.55	15.06	0.60	0.80
1620.00	69.50	0.58	15.74	0.63	0.83
2640.00	414.00	0.41	11.03	0.44	0.58
2760.00	47.50	0.40	10.75	0.43	0.57
2880.00	47.00	0.39	10.64	0.42	0.56
3000.00	48.00	0.40	10.87	0.43	0.57
3120.00	41.00	0.34	9.28	0.37	0.49
4500.00	468.00	0.34	9.21	0.37	0.49
4538.00	13.00	0.34	9.30	0.37	0.49
5520.00	398.00	0.41	11.01	0.44	0.58
5640.00	48.00	0.40	10.87	0.43	0.57
5760.00	44.00	0.37	9.96	0.40	0.53
5940.00	55.50	0.31	8.38	0.33	0.44
6000.00	20.00	0.33	9.06	0.36	0.48
7405.00	364.00	0.26	7.04	0.28	0.37
7435.00	6.00	0.20	5.43	0.22	0.29
8760.00	340.00	0.26	6.97	0.28	0.37
8820.00	17.50	0.29	7.92	0.32	0.42
9900.00	319.00	0.30	8.03	0.32	0.42
10020.00	41.50	0.35	9.40	0.37	0.50

RUN NO. VIII
MEMBRANE: PS2
PRESSURE: 100.00 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)
60.00	118.50	1.98	53.66	2.13	2.84
120.00	75.00	1.25	33.96	1.35	1.80
180.00	67.00	1.12	30.34	1.21	1.60
240.00	63.00	1.05	28.53	1.13	1.51
1260.00	412.00	0.40	10.97	0.44	0.58
1380.00	74.50	0.62	16.87	0.67	0.89
1500.00	70.50	0.59	15.96	0.63	0.84
1620.00	72.00	0.60	16.30	0.65	0.86
2640.00	450.00	0.44	11.99	0.48	0.63
2760.00	50.00	0.42	11.32	0.45	0.60
2880.00	50.50	0.42	11.43	0.45	0.60
3000.00	52.00	0.43	11.77	0.47	0.62
3120.00	43.50	0.36	9.85	0.39	0.52
4500.00	256.00	0.19	5.04	0.20	0.27
4538.00	6.50	0.17	4.65	0.18	0.25
5520.00	421.00	0.43	11.65	0.46	0.62
5640.00	51.00	0.43	11.55	0.46	0.61
5760.00	46.50	0.39	10.53	0.42	0.56
5940.00	59.50	0.33	8.98	0.36	0.47
6000.00	21.00	0.35	9.51	0.38	0.50
7405.00	108.00	0.08	2.09	0.08	0.11
7435.00	2.00	0.07	1.81	0.07	0.10
8760.00	362.00	0.27	7.42	0.30	0.39
8820.00	18.00	0.30	8.15	0.32	0.43
9900.00	338.00	0.31	8.50	0.34	0.45
10020.00	44.50	0.37	10.08	0.40	0.53

```
           RUN NO.  VIII
           MEMBRANE: PS3
           PRESSURE=  150.00 psi
           PRESSURE CORRECTION FACTOR 1.00
   TIME    PERMEATE PERMEATE   FLUX       FLUX       CORR
            VOLUME    RATE     RATE       RATE    FLUX RATE
   (MIN)    (MLS)  (MLS/MIN)(US GFD) (M3/M2·D)(M3/M2·D)
    60.00   102.00     1.70    46.19      1.84      1.84
   120.00    79.00     1.32    35.77      1.42      1.42
   180.00    72.00     1.20    32.60      1.30      1.30
   240.00    64.00     1.07    28.98      1.15      1.15
  1260.00   438.00     0.43    11.67      0.46      0.46
  1380.00    71.00     0.59    16.08      0.64      0.64
  1500.00    68.00     0.57    15.40      0.61      0.61
  1620.00    65.00     0.54    14.72      0.59      0.59
  2640.00   359.00     0.35     9.56      0.38      0.38
  2760.00    33.00     0.28     7.47      0.30      0.30
  2880.00    31.50     0.26     7.13      0.28      0.28
  3000.00    31.00     0.26     7.02      0.28      0.28
  3120.00    25.00     0.21     5.66      0.23      0.23
  4500.00   500.00     0.36     9.84      0.39      0.39
  4538.00    14.00     0.37    10.01      0.40      0.40
  5520.00   157.00     0.16     4.34      0.17      0.17
  5640.00    17.50     0.15     3.96      0.16      0.16
  5760.00    15.50     0.13     3.51      0.14      0.14
  5940.00    19.00     0.11     2.87      0.11      0.11
  6000.00     7.00     0.12     3.17      0.13      0.13
  7405.00   383.00     0.27     7.41      0.29      0.29
  7435.00     7.00     0.23     6.34      0.25      0.25
  8760.00    94.00     0.07     1.93      0.08      0.08
  8820.00     4.60     0.08     2.08      0.08      0.08
  9900.00    82.00     0.08     2.06      0.08      0.08
 10020.00    12.00     0.10     2.72      0.11      0.11
```

RUN NO. VIII
MEMBRANE: PS4
PRESSURE: 130.00 psi
PRESSURE CORRECTION FACTOR 1.13

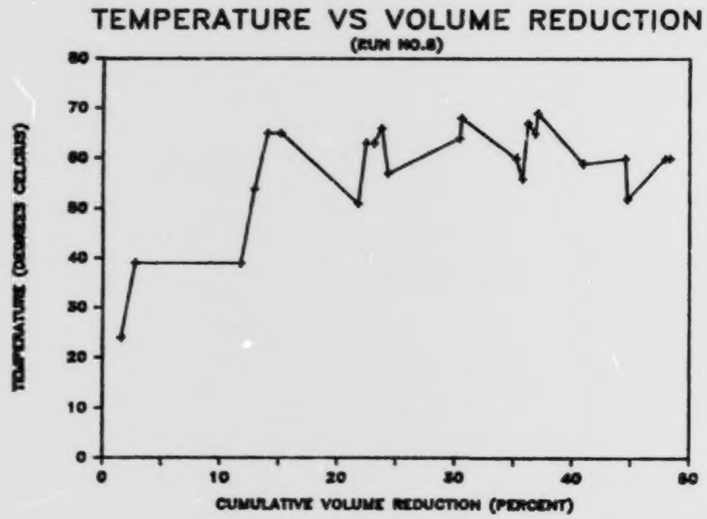
TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)
60.00	100.00	1.67	45.28	1.80	2.03
120.00	74.00	1.23	33.51	1.33	1.51
180.00	66.00	1.10	29.89	1.19	1.34
240.00	62.00	1.03	28.08	1.12	1.26
1260.00	410.00	0.40	10.92	0.43	0.49
1380.00	70.00	0.58	15.85	0.63	0.71
1500.00	66.50	0.55	15.06	0.60	0.68
1620.00	68.00	0.57	15.40	0.61	0.69
2640.00	428.00	0.42	29.89	0.45	0.51
2760.00	44.00	0.37	24.91	0.40	0.45
2880.00	40.50	0.34	24.45	0.36	0.41
3000.00	43.00	0.36	22.64	0.39	0.44
3120.00	37.00	0.31	21.74	0.33	0.38
4500.00	412.00	0.30	20.83	0.32	0.36
4538.00	10.00	0.26	19.47	0.28	0.32
5520.00	333.00	0.34	18.57	0.37	0.41
5640.00	38.50	0.32	8.72	0.35	0.39
5760.00	35.00	0.29	7.92	0.32	0.36
5940.00	44.00	0.24	6.64	0.26	0.30
6000.00	11.00	0.18	4.98	0.20	0.22
7405.00	174.00	0.12	3.36	0.13	0.15
7435.00	3.00	0.10	2.72	0.11	0.12
8760.00	266.00	0.20	5.45	0.22	0.25
8820.00	13.00	0.22	5.89	0.23	0.26
9900.00	241.00	0.22	6.06	0.24	0.27
10020.00	31.50	0.26	7.13	0.28	0.32

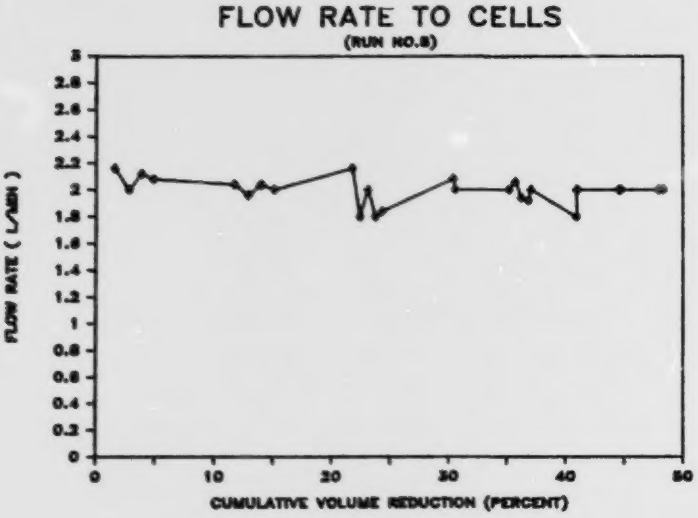
RUN NO. VIII
MEMBRANE: PS5
PRESSURE: 140.00 psi
PRESSURE CORRECTION FACTOR 1.07

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)
60.00	89.00	1.48	40.30	1.60	1.71
120.00	76.50	1.28	34.64	1.38	1.47
180.00	71.50	1.19	32.38	1.29	1.38
240.00	66.00	1.10	29.89	1.19	1.27
1260.00	477.00	0.47	12.71	0.51	0.54
1380.00	77.00	0.64	17.43	0.69	0.74
1500.00	72.50	0.60	16.42	0.65	0.70
1620.00	68.00	0.57	15.40	0.61	0.65
2640.00	376.00	0.37	10.02	0.40	0.43
2760.00	33.00	0.28	7.47	0.30	0.32
2880.00	32.00	0.27	7.25	0.29	0.31
3000.00	31.00	0.26	7.02	0.28	0.30
3120.00	24.50	0.20	5.55	0.22	0.24
4500.00	262.00	0.19	5.16	0.21	0.22
4538.00	7.50	0.20	5.36	0.21	0.23
5520.00	193.00	0.20	5.34	0.21	0.23
5640.00	23.00	0.19	5.21	0.21	0.22
5760.00	20.50	0.17	4.64	0.18	0.20
5940.00	25.50	0.14	3.85	0.15	0.16
6000.00	9.50	0.16	4.30	0.17	0.18
7405.00	146.00	0.10	2.82	0.11	0.12
7435.00	2.50	0.08	2.26	0.09	0.10
8760.00	133.00	0.10	2.73	0.11	0.12
8820.00	7.00	0.12	3.17	0.13	0.13
9900.00	117.00	0.11	2.94	0.12	0.13
10020.00	7.00	0.06	1.58	0.06	0.07

RUN NO. VIII
MEMBRANE: PS6
PRESSURE: 150.00 psi
PRESSURE CORRECTION FACTOR 1.00

TIME (MIN)	PERMEATE VOLUME (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)
60.00	101.00	1.68	45.74	1.82	1.82
120.00	81.00	1.35	36.68	1.46	1.46
180.00	74.00	1.23	33.51	1.33	1.33
240.00	69.50	1.16	31.47	1.25	1.25
1260.00	493.00	0.48	13.13	0.52	0.52
1380.00	78.00	0.65	17.66	0.70	0.70
1500.00	76.00	0.63	17.21	0.68	0.68
1620.00	76.00	0.63	17.21	0.68	0.68
2640.00	492.00	0.48	13.11	0.52	0.52
2760.00	48.50	0.40	10.98	0.44	0.44
2880.00	50.00	0.42	11.32	0.45	0.45
3000.00	47.00	0.39	10.64	0.42	0.42
3120.00	39.00	0.33	8.83	0.35	0.35
4500.00	427.00	0.31	8.41	0.33	0.33
4538.00	11.00	0.29	7.87	0.31	0.31
5520.00	252.00	0.26	6.97	0.28	0.28
5640.00	26.50	0.22	6.00	0.24	0.24
5760.00	23.50	0.20	5.32	0.21	0.21
5940.00	29.00	0.16	4.38	0.17	0.17
6000.00	11.00	0.18	4.98	0.20	0.20
7405.00	288.00	0.20	5.57	0.22	0.22
7435.00	5.00	0.17	4.53	0.18	0.18
8760.00	154.00	0.12	3.16	0.13	0.13
8820.00	7.50	0.13	3.40	0.14	0.14
9900.00	131.00	0.12	3.30	0.13	0.13
10020.00	18.00	0.15	4.08	0.16	0.16


TEMPERATURE VS VOLUME REDUCTION
(RUN NO.8)


FLOW RATE TO CELLS
(RUN NO.8)

```
          CORRELATION AND MULTIPLE LINEAR REGRESSION RESULTS

                  CHP FLUX RATES AND CONDITIONS


 1              .55 0 -1
 2              .79 0 0
 3              .48 -1 -1
 4              .45 -1 0
 5              .55 1 -1
 6              .99 1 0
 7              .81 0 1
                   CORRELATION OF CHP DATA

COL A     MEAN          STD DEV     COL B  COVARIANCE    CORRELATION
-----  -----------   -----------    -----  -----------   -----------

 1      .66           .2037156       2      8.714287E-02  .6112233
                                     3      7.857143E-02  .5952591
 2      0             .8164966       3      0             0
 3     -.2857143      .7559289

        # OF ROWS =   7


                  MULTIPLE LINEAR REGRESSION

TERM           COEFFICIENT    T-TEST
----           -----------   ----------

B 0            .7058333      0
B 1            .1525         2.343627
B 2            .1604167      2.282416
                      SUM SQ       DEG FR      MEAN SQ
                      ------       ------      -------

DUE TO REGRESSION     .1812542       2         9.062709E-02
ABOUT REGRESSION      6.774584E-02   4         1.693646E-02
TOTAL                 .249           6         .0415

R-SQUARED       .7279284
F-TEST          5.351006
```

```
                PS FLUX RATES AND CONDITIONS


1              .7 0 -1
2              .78 0 0
3              .18 -1 -1
4              .88 -1 0
5              .57 1 -1
6              .99 1 0
7              1.10 0 1
                    CORRELATION FOR PS DATA

COL A     MEAN          STD DEV       COL B   COVARIANCE     CORRELATION
-----  -----------    -----------     -----  -----------    -----------

 1      .7428571       .3047795        2       7.142858E-02   .3348719
                                       3       .1622449       .8215823
 2      0              .8164966        3       0              0
 3     -.2857143       .7559289

          # OF ROWS =  7


                     MULTIPLE LINEAR REGRESSION FOR PS DATA

TERM              COEFFICIENT      T-TEST
----              -----------    ----------

B 0               .8375           0
B 1               .125            1.45164
B 2               .33125          3.561488
                          SUM SQ       DEG FR       MEAN SQ
                          ------       ------       -------

DUE TO REGRESSION        .4387054        2          .2193527
ABOUT REGRESSION         .1186375        4          2.965938E-02
TOTAL                    .5573429        6          9.289048E-02

R-SQUARED         .7871373
F-TEST            7.395729
```

```
                 TFC FLUX RATES AND CONDITIONS


 1             .63 0 -1
 2             .75 0 0
 3             .39 -1 -1
 4             .35 -1 0
 5             .5 1 -1
 6             .9 1 0
 7             1.05 0 1
                  CORRELATION OF TFC DATA

COL A     MEAN           STD DEV      COL B   COVARIANCE      CORRELATION
-----  -----------    -----------     -----   -----------     -----------

 1      .6528571       .2623429        2       9.428571E-02    .5135338
                                       3       .1193878       .7023545
 2     0               .8164966        3       0              0
 3     -.2857143       .7559289

        # OF ROWS =  7


                   MULTIPLE LINEAR REGRESSION OF TFC DATA

TERM             COEFFICIENT     T-TEST
----             -----------    ----------

B 0              .7225          0
B 1              .165           2.08359
B 2              .24375         2.849703

                          SUM SQ         DEG FR       MEAN SQ
                          ------         ------       -------

DUE TO REGRESSION        .3126053          2           .1563027
ABOUT REGRESSION         .1003375          4           2.508438E-02
TOTAL                    .4129428          6           6.882381E-02

R-SQUARED        .7570184
F-TEST           6.231076
```

```
                CELLULOSE FLUX RATES AND CONDITIONS


1            .22 0 -1
2            .25 0 0
3            .26 -1 -1
4            .28 -1 0
5            .21 1 -1
6            .38 1 0
                 CORRELATION OF CELLULOSE DATA

COL A     MEAN          STD DEV       COL B   COVARIANCE     CORRELATION
-----  -----------    -----------     -----  -----------    -----------

  1     .2666667      6.120991E-02    2       8.333336E-03   .1826558
                                      3       1.833334E-02   .6562065
  2    0               .8944272       3      0               0
  3    -.5             .5477226

        # OF ROWS =  6


                  MULTIPLE LINEAR REGRESSION OF CELLULOSE DATA

TERM            COEFFICIENT      T-TEST
----            -----------    ----------

B 0             .3033334        0
B 1             .0125          .4321134
B 2             7.333334E-02   1.552404
                         SUM SQ        DEG FR        MEAN SQ
                         ------        ------        -------

DUE TO REGRESSION       8.691668E-03    2          4.345834E-03
ABOUT REGRESSION        1.004167E-02    3          3.347222E-03
TOTAL                   1.873333E-02    5          3.746667E-03

R-SQUARED       .463968
F-TEST          1.298341
```

```
                    CA FLUX RATES AND CONDITIONS



1            .59 0 -1
2            .74 0 0
3            .51 -1 -1
4            .88 -1 0
5            .76 1 -1
6            1.07 1 0
7            1.03 0 1

                    CORRELATION OF CA DATA

COL A    MEAN          STD DEV       COL B   COVARIANCE      CORRELATION
-----  -----------   -----------     -----   -----------     -----------

 1      .7971429      .2103737        2      6.285715E-02    .426929
                                      3      .1091837        .8009993
 2      0             .8164966        3      0               0
 3     -.2857143      .7559289

          # OF ROWS =  7


                 (D5): MULTIPLE LINEAR REGRESSION OF CA DATA

TERM             COEFFICIENT      T-TEST
----             -----------    ----------

B 0              .8608334       0
B 1              .11            2.034535
B 2              .2229167       3.81717

                      SUM SQ        DEG FR       MEAN SQ
                      ------        ------       -------

DUE TO REGRESSION     .2187721        2           .109386
ABOUT REGRESSION      4.677086E-02    4           1.169271E-02
TOTAL                 .2655429        6           4.425715E-02

R-SQUARED        .823867
F-TEST           9.355059
```

```
                PVDF FLUX RATES AND CONDITIONS


 1             .59 0 -1
 2             .71 0 0
 3             .39 -1 -1
 4             .65 -1 0
 5             .45 1 -1
 6             .77 1 0
 7             .89 0 1
                 CORRELATION DATA FOR PVDF MEMBRANE

 COL A    MEAN          STD DEV      COL B   COVARIANCE     CORRELATION
 -----  -----------   -----------    -----  -----------    -----------

  1      .6357143      .1757705        2     2.571429E-02   .2090359
                                       3     .1044898       .9174739
  2      0             .8164966        3     0              0
  3     -.2857143      .7559289

         # OF ROWS =  7


                  MULTIPLE LINEAR REGRESSION FOR PVDF

 TERM           COEFFICIENT     T-TEST
 ----           -----------    ----------

 B 0            .6966666       0
 B 1            .045           1.235274
 B 2            .2133334       5.421709
                            SUM SQ       DEG FR      MEAN SQ
                            ------       ------      -------

 DUE TO REGRESSION       .1641381         2          8.206905E-02
 ABOUT REGRESSION        2.123334E-02     4          5.308334E-03
 TOTAL                   .1853714         6          3.089524E-02

 R-SQUARED      .8854552
 F-TEST         15.46042
```

FLUX RATE CHANGE WITH pH

MEMBRANE	MEAN FLUX RATE AT LOW pH (5.5)	MEAN FLUX RATE AT NORMAL pH (7.5-8.0)	MEAN FLUX RATE AT HIGH pH (9.5)
CHP	0.69	0.91	1.02
PS	0.58	1.09	1.10
TFC	0.57	0.87	0.95
CELLULOSE	0.29	0.29	0.31
CA	1.02	0.99	1.29
PVDF	0.80	0.98	0.91

APPENDIX E

RESULTS OF CONCENTRATION AND STABILITY TEST

BATCH CONCENTRATION EXPERIMENT
% CUMULATIVE VOLUME REDUCTION
OCTOBER 3 17 1984
VOLUME OF FEED 30 L

TIME	CHP	TFC	PS	TFC	PS	CHP	PVDF LRG CELLS	TOTAL	CUM TOT	VOLUME REDUC.	CUM VOLUME REDUC.	FEED RATE	FEED TEMP
(HRS)	PERM VOL (MLS)	PERM VOL (MLS)	PERM VOL (MLS)	PERM VOL (MLS)	PERM VOL (MLS)	PERM VOL (MLS)	PERM VOL (MLS)	PERM VOL (MLS)	PERM VOL (MLS)	%	%	L/MIN	C
1.00	189.00	78.00	117.00	62.00	121.00	115.00		682.00	682.00	2.27	2.27	2.16	48
3.67	350.00	280.00	270.00	255.00	304.00	300.00		1759.00	2441.00	5.86	8.14		
17.00	750.00	655.00	574.00	527.00	569.00	444.00		3519.00	5960.00	11.73	19.87		58
18.00	42.00	29.00	18.00	15.00	16.00	10.00	3237.00	3367.00	9327.00	11.22	31.09	1.96	63
19.00	38.00	29.00	19.00	17.00	18.00	11.00	150.00	282.00	9609.00	0.94	32.03	2.00	64
20.00	46.00	31.00	20.00	19.00	19.00	12.00	1490.00	1637.00	11246.00	5.46	37.49	2.32	66
22.00	79.00	59.00	35.00	36.00	36.00	25.00	2741.00	3011.00	14257.00	10.04	47.52	2.16	63
23.00	34.00	28.00	17.00	18.00	18.00	13.00	1275.00	1403.00	15660.00	4.68	52.20	2.16	62
24.00	34.00	28.00	16.00	17.00	18.00	13.00	1260.00	1386.00	17046.00	4.62	56.82	2.20	60
25.00	35.00	26.00	16.00	17.00	18.00	13.00	910.00	1035.00	18081.00	3.45	60.27		62
41.00	496.00	426.00	246.00	290.00	337.00	259.00		2054.00	20135.00	6.85	67.12	1.80	66
42.00	30.00	29.00	17.00	19.00	24.00	18.00	1375.00	1512.00	21647.00	5.04	72.16	2.20	62
43.00	29.00	30.00	17.00	20.00	26.00	20.00	1390.00	1532.00	23179.00	5.11	77.26	1.96	64
44.00	27.00	28.00	16.00	20.00	26.00	21.00	1330.00	1468.00	24647.00	4.89	82.16	1.60	65
46.00	30.00	30.00	16.00	20.00	29.00	23.00	1845.00	1993.00	26640.00	6.64	88.80	1.60	55
47.00	6.00	6.00	3.00	4.00	5.00	5.00	360.00	389.00	27029.00	1.30	90.10	1.40	

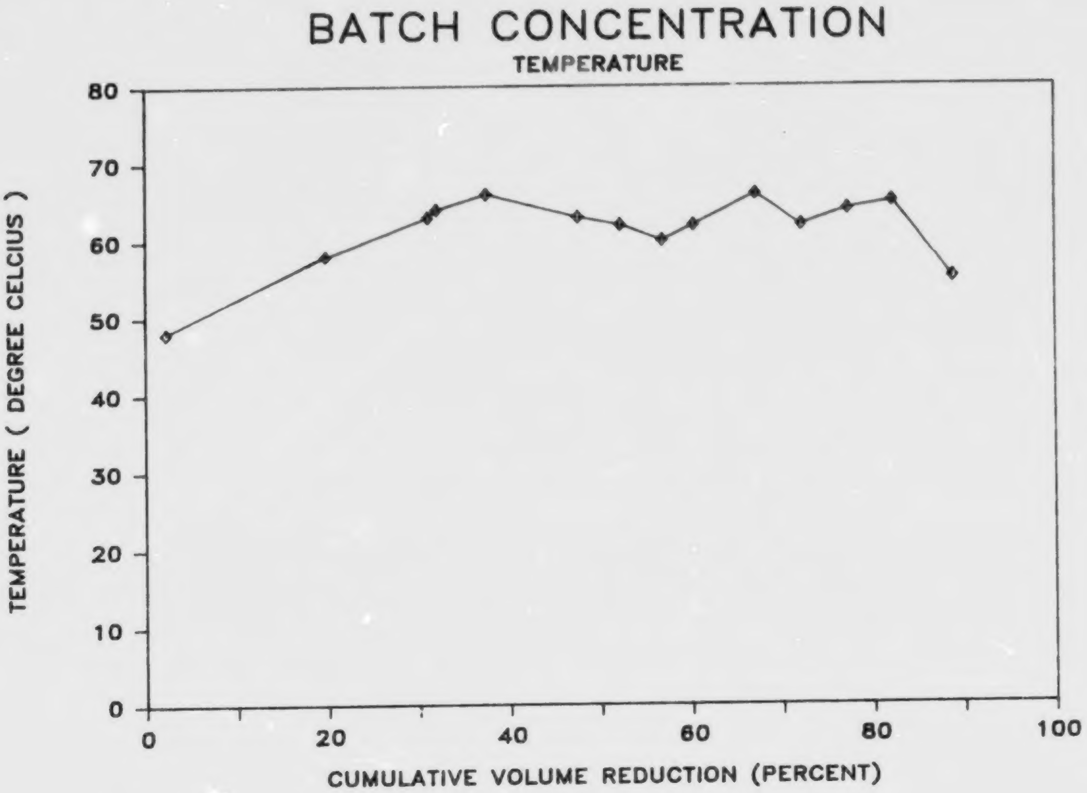
```
                                   CHP CELL1
                                   PRESSURE: 150 psi
                                   PRESSURE CORRECTION FACTOR 1.00
  TIME      PERM      PERM      FLUX      FLUX      CORR       CUM
           VOLUME     RATE      RATE      RATE    FLUX RATE  VOL RED
  (MIN)    (MLS)  (MLS/MIN)(US GFD) (M3/M2·D)(M3/M2·D)   (%)
      0
   60.0   189.00     3.15     85.59      3.40      3.40      2.27
  220.2   350.00     2.18     59.36      2.36      2.36      8.14
 1020.0   750.00     0.94     25.48      1.01      1.01     19.87
 1080.0    42.00     0.70     19.02      0.76      0.76     31.09
 1140.0    38.00     0.63     17.21      0.68      0.68     32.03
 1200.0    46.00     0.77     20.83      0.83      0.83     37.49
 1320.0    79.00     0.66     17.89      0.71      0.71     47.52
 1380.0    34.00     0.57     15.40      0.61      0.61     52.20
 1440.0    34.00     0.57     15.40      0.61      0.61     56.82
 1500.0    35.00     0.58     15.85      0.63      0.63     60.27
 2460.0   496.00     0.52     14.04      0.56      0.56     67.12
 2520.0    30.00     0.50     13.59      0.54      0.54     72.16
 2580.0    29.00     0.48     13.13      0.52      0.52     77.26
 2640.0    27.00     0.45     12.23      0.49      0.49     82.16
 2760.0    30.00     0.25      6.79      0.27      0.27     88.80
 2820.0     6.00     0.10      2.72      0.11      0.11     90.10
```

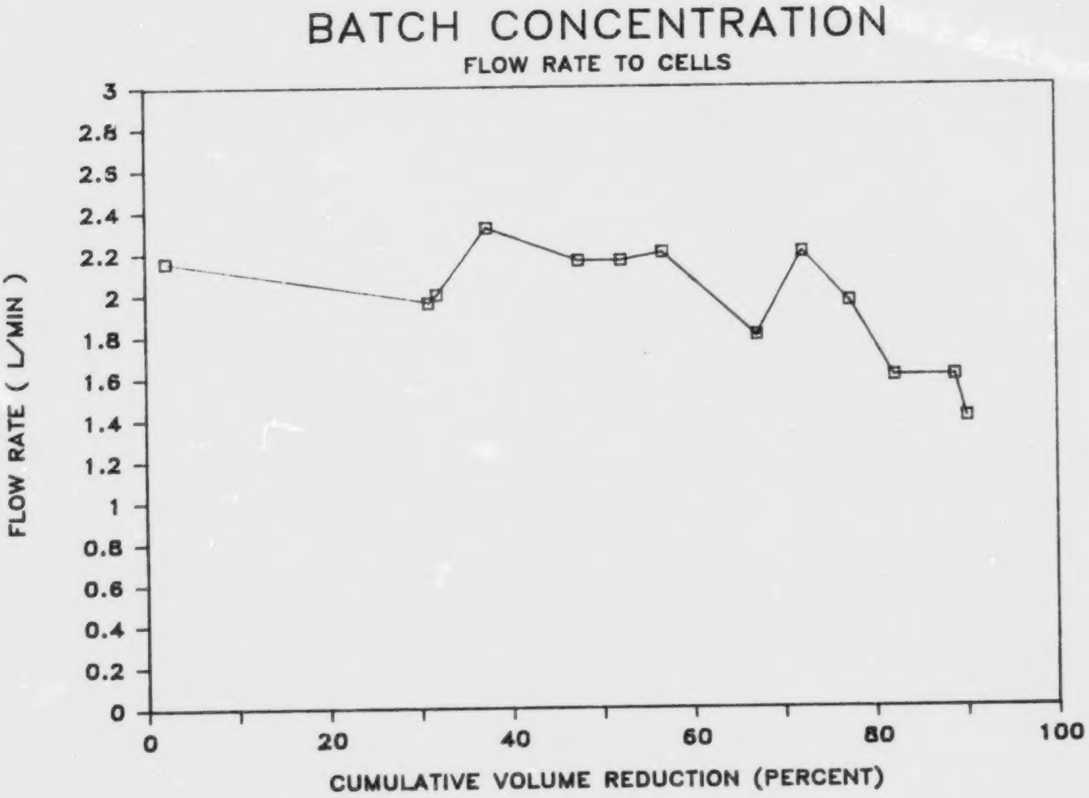
TFC CELL 2
PRESSURE: 150 psi
PRESSURE CORRECTION FACTOR 1.00

TIME (MIN)	PERM VOLUME (MLS)	PERM RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	CUM VOL RED (%)
0						
60.0	78.00	1.30	35.32	1.40	1.40	2.27
220.2	280.00	1.75	47.49	1.89	1.89	8.14
1020.0	655.00	0.82	22.25	0.88	0.88	19.87
1080.0	29.00	0.48	13.13	0.52	0.52	31.09
1140.0	29.00	0.48	13.13	0.52	0.52	32.03
1200.0	31.00	0.52	14.04	0.56	0.56	37.49
1320.0	59.00	0.49	13.36	0.53	0.53	47.52
1380.0	28.00	0.47	12.68	0.50	0.50	52.20
1440.0	28.00	0.47	12.68	0.50	0.50	56.82
1500.0	26.00	0.43	11.77	0.47	0.47	60.27
2460.0	426.00	0.44	12.06	0.48	0.48	67.12
2520.0	29.00	0.48	13.13	0.52	0.52	72.16
2580.0	30.00	0.50	13.59	0.54	0.54	77.26
2640.0	28.00	0.47	12.68	0.50	0.50	82.16
2760.0	30.00	0.25	6.79	0.27	0.27	88.80
2820.0	6.00	0.10	2.72	0.11	0.11	90.10

PS CELL 3
PRESSURE: 140 psi
PRESSURE CORRECTION FACTOR 1.07

TIME (MIN)	PERM VOLUME (MLS)	PERM RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2·D)	CORR FLUX RATE (M3/M2·D)	CUM VOL RED (%)
0						
60.0	117.00	1.950	52.982	2.106	2.253	2.27
220.2	270.00	1.685	45.792	1.820	1.948	8.14
1020.0	574.00	0.718	19.499	0.775	0.829	19.87
1080.0	18.00	0.300	8.151	0.324	0.347	31.09
1140.0	19.00	0.317	8.604	0.342	0.366	32.03
1200.0	20.00	0.333	9.057	0.360	0.385	37.49
1320.0	35.00	0.292	7.925	0.315	0.337	47.52
1380.0	17.00	0.283	7.698	0.306	0.327	52.20
1440.0	16.00	0.267	7.245	0.288	0.308	56.82
1500.0	16.00	0.267	7.245	0.288	0.308	60.27
2460.0	246.00	0.256	6.962	0.277	0.296	67.12
2520.0	17.00	0.283	7.698	0.306	0.327	72.16
2580.0	17.00	0.283	7.698	0.306	0.327	77.26
2640.0	16.00	0.267	7.245	0.288	0.308	82.16
2760.0	16.00	0.133	3.623	0.144	0.154	88.80
2820.0	3.00	0.050	1.359	0.054	0.058	90.10


BATCH CONCENTRATION
TEMPERATURE
TEMPERATURE (DEGREE CELCIUS)
80
70
60
50
40
30
20
10
0
0
20
40
60
80
100
CUMULATIVE VOLUME REDUCTION (PERCENT)


BATCH CONCENTRATION
FLOW RATE TO CELLS
FLOW RATE (L/MIN)
3
2.8
2.6
2.4
2.2
2
1.8
1.6
1.4
1.2
1
0.8
0.6
0.4
0.2
0
0
20
40
60
80
100
CUMULATIVE VOLUME REDUCTION (PERCENT)

STABILITY EXPERIMENT
% CUMULATIVE VOLUME REDUCTION
OCTOBER 3 17 1984

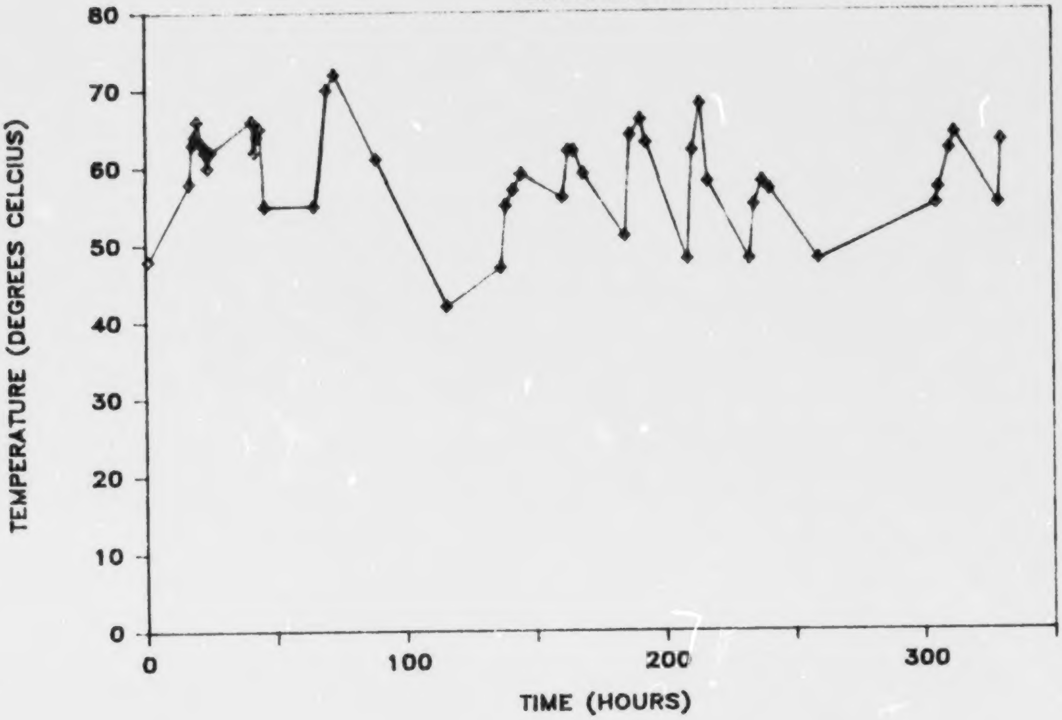
TIME	CHP	TFC	PS	TFC	VOLUME OF FEED PS	CHP	30 L PVDF LRG CELLS	TOTAL	CUM TOT	VOLUME REDUC.	CUM VOLUME REDUC.	FEED RATE	FEED TEMP
	PERM VOL	PERM VOL	PERM VOL	PERM VOL	PERM VOL	PERM VOL	PERM VOL	PERM VOL	PERM VOL				
(HRS)	(MLS)	(MLS)	(MLS)	(MLS)	(MLS)	(MLS)	(MLS)	(MLS)	(MLS)	%	%	L/MIN	C
1.00	189.00	78.00	117.00	62.00	121.00	115.00		682.00	682.00	2.27	2.27	2.16	48
3.67	350.00	280.00	270.00	255.00	304.00	300.00		1759.00	2441.00	5.86	8.14		
17.00	750.00	655.00	574.00	527.00	569.00	444.00		3519.00	5960.00	11.73	19.87		58
18.00	42.00	29.00	18.00	15.00	16.00	10.00	3237.00	3367.00	9327.00	11.22	31.09	1.96	63
19.00	38.00	29.00	19.00	17.00	18.00	11.00	150.00	282.00	9609.00	0.94	32.03	2.00	64
20.00	46.00	31.00	20.00	19.00	19.00	12.00	1490.00	1637.00	11246.00	5.46	37.49	2.32	66
22.00	79.00	59.00	35.00	36.00	36.00	25.00	2741.00	3011.00	14257.00	10.04	47.52	2.16	63
23.00	34.00	28.00	17.00	18.00	18.00	13.00	1275.00	1403.00	15660.00	4.68	52.20	2.16	62
24.00	34.00	28.00	16.00	17.00	18.00	13.00	1260.00	1386.00	17046.00	4.62	56.82	2.20	60
25.00	35.00	26.00	16.00	17.00	18.00	13.00	910.00	1035.00	18081.00	3.45	60.27		62
41.00	496.00	426.00	246.00	290.00	337.00	259.00		2054.00	20135.00	6.85	67.12	1.80	66
42.00	30.00	29.00	17.00	19.00	24.00	18.00	1375.00	1512.00	21647.00	5.04	72.16	2.20	62
43.00	29.00	30.00	17.00	20.00	26.00	20.00	1390.00	1532.00	23179.00	5.11	77.26	1.96	64
44.00	27.00	28.00	16.00	20.00	26.00	21.00	1330.00	1468.00	24647.00	4.89	82.16	1.60	65
46.00	30.00	30.00	16.00	20.00	29.00	23.00	1845.00	1993.00	26640.00	6.64	88.80	1.60	55
47.00	6.00	6.00	3.00	4.00	5.00	5.00	360.00	389.00	27029.00	1.30	90.10	1.40	
49.00	6.00	5.50	3.00	3.00	5.00	3.50					90.10	1.20	
65.00	65.50	54.50	17.50	22.00	44.00	22.00					90.10	1.20	55
67.00	24.00	29.00	6.50	19.50	30.50	30.00					90.10		
70.00	47.50	59.00	11.50	36.00	59.00	61.50					90.10	1.60	70
73.00	38.50	46.50	9.00	29.00	48.00	51.00					90.10	1.72	72
88.92	142.00	162.00	30.00	107.00	185.00	202.00					90.10	1.80	61
96.48	93.00	76.00	15.00	53.00	58.00	109.00					90.10		
115.98	124.00	159.00	30.00	114.00	203.00	242.00					90.10		42
137.00	91.00	126.00	24.00	103.00	141.00	213.00					90.10	2.32	47

CON'T

TIME (HRS)	CHP PERM VOL (MLS)	TFC PERM VOL (MLS)	PS PERM VOL (MLS)	TFC PERM VOL (MLS)	PS PERM VOL (MLS)	CHP PERM VOL (MLS)	PVDF LRG CELLS PERM VOL (MLS)	TOTAL PERM VOL (MLS)	CUM TOT PERM VOL (MLS)	VOLUME REDUC. %	CUM VOLUME REDUC. %	FEED RATE L/MIN	FEED TEMP C
139.00	11.50	16.00	3.00	13.00	16.00	24.50					90.10	1.60	55
142.00	17.50	24.00	4.00	19.00	24.00	36.50					90.10	2.16	57
145.00	17.00	23.50	4.00	19.00	24.00	37.00					90.10	2.04	59
161.00	68.50	90.50	14.00	70.00	89.00	145.00					90.10	2.06	56
163.00	13.00	17.50	3.00	13.00	14.50	22.50					90.10	1.60	62
163.25	3.50	4.00	0.50	2.50	2.50	2.50					90.10	2.64	
165.00	10.50	15.00	2.50	12.00	14.00	25.00					90.10		62
169.00	19.00	28.00	4.50	23.00	27.00	52.00					90.10	2.12	59
185.00	69.50	99.50	13.00	74.50	75.50	158.00					90.10	1.72	51
187.00	13.00	19.00	2.50	14.00	11.00	24.00					90.10	1.52	64
191.00	24.50	38.50	4.00	26.00	19.00	50.00					90.10	2.08	66
193.00	9.00	15.00	1.00	12.00	8.00	23.50					90.10	2.40	63
209.00	45.50	73.50	6.00	55.00	36.00	117.00					90.10	2.32	48
211.00	8.50	15.00	1.50	11.50	7.00	20.50					90.10	1.80	62
214.00	15.00	28.00	2.50	19.50	13.00	42.00					90.10	1.92	68
217.00	12.50	23.00	2.00	17.00	9.50	30.50					90.10		58
233.00	45.50	94.00	7.00	69.50	37.50	120.00					90.10		48
235.00	6.50	13.00	1.00	10.00	5.00	15.00					90.10		55
238.00	10.50	20.50	2.00	16.00	8.00	23.50					90.10		58
241.00	9.50	19.00	1.00	15.00	6.50	22.00					90.10		57
259.50	37.50	86.00	4.50	50.00	23.00	76.00					90.10		48
286.00	37.50	74.00	3.50	56.00	21.00	84.00					90.10		
305.00	113.00	204.00	28.00	151.00	39.50	179.00					90.10		55
306.00	3.50	7.00	0.50	6.00	1.00	8.50					90.10		57
310.00	15.00	29.50	2.50	25.00	5.50	34.50					90.10		62
312.00	7.50	14.50	1.00	13.00	2.50	18.00					90.10		64
329.00	43.00	86.00	3.50	65.00	14.00	103.00					90.10		55
330.00	3.50	7.00	0.20	6.00	1.00	8.00					90.10		63

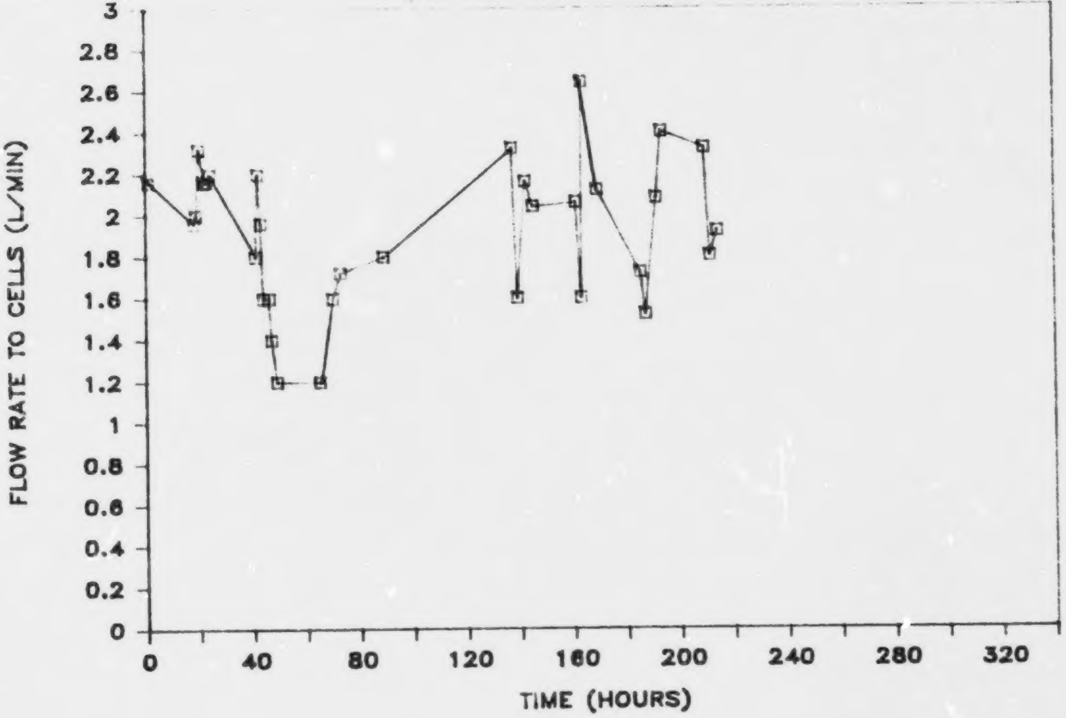
STABILITY EXPERIMENT

TEMPERATURE



STABILITY EXPERIMENT

(TEMPERATURE 65C, pH AS IS)



STABILITY EXPERIMENT
TFC CELL 4
PRESSURE: 130 psi
PRESSURE CORRECTION FACTOR 1.13

TIME (MIN)	PERM RATE (MLS)	PERM RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2 D)	PRESS CORR (M3/M2 D)	CUM VOL RED (%)
0						
60.0	62.00	1.03	28.08	1.12	1.26	2.27
220.2	255.00	1.59	43.25	1.72	1.94	8.14
1020.0	527.00	0.66	17.90	0.71	0.80	19.87
1080.0	15.00	0.25	6.79	0.27	0.31	31.09
1140.0	17.00	0.28	7.70	0.31	0.35	32.03
1200.0	19.00	0.32	8.60	0.34	0.39	37.49
1320.0	36.00	0.30	8.15	0.32	0.37	47.52
1380.0	18.00	0.30	8.15	0.32	0.37	52.20
1440.0	17.00	0.28	7.70	0.31	0.35	56.82
1500.0	17.00	0.28	7.70	0.31	0.35	60.27
2460.0	290.00	0.30	8.21	0.33	0.37	67.12
2520.0	19.00	0.32	8.60	0.34	0.39	72.16
2580.0	20.00	0.33	9.06	0.36	0.41	77.26
2640.0	20.00	0.33	9.06	0.36	0.41	82.16
2760.0	20.00	0.17	4.53	0.18	0.20	88.80
2820.0	4.00	0.07	1.81	0.07	0.08	90.10
2940.0	3.00	0.03	0.68	0.03	0.03	90.10
3900.0	22.00	0.02	0.62	0.02	0.03	90.10
4020.0	19.50	0.16	4.42	0.18	0.20	90.10
4200.0	36.00	0.20	5.43	0.22	0.24	90.10
4380.0	29.00	0.16	4.38	0.17	0.20	90.10
5335.2	107.00	0.11	3.04	0.12	0.14	90.10
5788.8	53.00	0.12	3.17	0.13	0.14	90.10
6958.8	114.00	0.10	2.65	0.11	0.12	90.10
8220.0	103.00	0.08	2.22	0.09	0.10	90.10
8340.0	13.00	0.11	2.94	0.12	0.13	90.10
8520.0	19.00	0.11	2.87	0.11	0.13	90.10
8700.0	19.00	0.11	2.87	0.11	0.13	90.10
9660.0	70.00	0.07	1.98	0.08	0.09	90.10
9780.0	13.00	0.11	2.94	0.12	0.13	90.10
9795.0	2.50	0.17	4.53	0.18	0.20	90.10
9900.0	12.00	0.11	3.11	0.12	0.14	90.10
10140.0	23.00	0.10	2.60	0.10	0.12	90.10
11100.0	74.50	0.08	2.11	0.08	0.09	90.10
11220.0	14.00	0.12	3.17	0.13	0.14	90.10
11460.0	26.00	0.11	2.94	0.12	0.13	90.10
11580.0	12.00	0.10	2.72	0.11	0.12	90.10
12540.0	55.00	0.06	1.56	0.06	0.07	90.10
12660.0	11.50	0.10	2.60	0.10	0.12	90.10
12840.0	19.50	0.11	2.94	0.12	0.13	90.10
13020.0	17.00	0.09	2.57	0.10	0.12	90.10
13980.0	69.50	0.07	1.97	0.08	0.09	90.10
14100.0	10.00	0.08	2.26	0.09	0.10	90.10
14280.0	16.00	0.09	2.42	0.10	0.11	90.10
14460.0	15.00	0.08	2.26	0.09	0.10	90.10
15570.0	50.00	0.05	1.22	0.05	0.05	90.10
17160.0	56.00	0.04	0.96	0.04	0.04	90.10
18300.0	151.00	0.13	3.60	0.14	0.16	90.10
18360.0	6.00	0.10	2.72	0.11	0.12	90.10
18600.0	25.00	0.10	2.83	0.11	0.13	90.10
18720.0	13.00	0.11	2.94	0.12	0.13	90.10
19740.0	65.00	0.06	1.73	0.07	0.08	90.10
19800.0	6.00	0.10	2.72	0.11	0.12	90.10

STABILITY EXPERIMENT
PS CELL 5
PRESSURE: 100 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERM RATE (MLS)	PERM RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2 D)	PRESS CORR (M3/M2 D)	CUM VOL RED (%)
0						
60.0	121.00	2.02	54.79	2.18	2.90	2.27
220.2	304.00	1.90	51.56	2.05	2.73	8.14
1020.0	569.00	0.71	19.33	0.77	1.02	19.87
1080.0	16.00	0.27	7.25	0.29	0.38	31.09
1140.0	18.00	0.30	8.15	0.32	0.43	32.03
1200.0	19.00	0.32	8.60	0.34	0.45	37.49
1320.0	36.00	0.30	8.15	0.32	0.43	47.52
1380.0	18.00	0.30	8.15	0.32	0.43	52.20
1440.0	18.00	0.30	8.15	0.32	0.43	56.82
1500.0	18.00	0.30	8.15	0.32	0.43	60.27
2460.0	337.00	0.35	9.54	0.38	0.50	67.12
2520.0	24.00	0.40	10.87	0.43	0.57	72.16
2580.0	26.00	0.43	11.77	0.47	0.62	77.26
2640.0	26.00	0.43	11.77	0.47	0.62	82.16
2760.0	29.00	0.24	6.57	0.26	0.35	88.80
2820.0	5.00	0.08	2.26	0.09	0.12	90.10
2940.0	5.00	0.04	1.13	0.05	0.06	90.10
3900.0	44.00	0.05	1.25	0.05	0.07	90.10
4020.0	30.50	0.25	6.91	0.27	0.37	90.10
4200.0	59.00	0.33	8.91	0.35	0.47	90.10
4380.0	48.00	0.27	7.25	0.29	0.38	90.10
5335.2	185.00	0.19	5.26	0.21	0.28	90.10
5788.8	58.00	0.13	3.47	0.14	0.18	90.10
6958.3	203.00	0.17	4.71	0.19	0.25	90.10
8220.0	141.00	0.11	3.04	0.12	0.16	90.10
8340.0	16.00	0.13	3.62	0.14	0.19	90.10
8520.0	24.00	0.13	3.62	0.14	0.19	90.10
8700.0	24.00	0.13	3.62	0.14	0.19	90.10
9660.0	89.00	0.09	2.52	0.10	0.13	90.10
9780.0	14.50	0.12	3.28	0.13	0.17	90.10
9795.0	2.50	0.17	4.53	0.18	0.24	90.10
9900.0	14.00	0.13	3.62	0.14	0.19	90.10
10140.0	27.00	0.11	3.06	0.12	0.16	90.10
11100.0	75.50	0.08	2.14	0.08	0.11	90.10
11220.0	11.00	0.09	2.49	0.10	0.13	90.10
11460.0	19.00	0.08	2.15	0.09	0.11	90.10
11580.0	8.00	0.07	1.81	0.07	0.10	90.10
12540.0	36.00	0.04	1.02	0.04	0.05	90.10
12660.0	7.00	0.06	1.58	0.06	0.08	90.10
12840.0	13.00	0.07	1.96	0.08	0.10	90.10
13020.0	9.50	0.05	1.43	0.06	0.08	90.10
13980.0	37.50	0.04	1.06	0.04	0.06	90.10
14100.0	5.00	0.04	1.13	0.05	0.06	90.10
14280.0	8.00	0.04	1.21	0.05	0.06	90.10
14460.0	6.50	0.04	0.98	0.04	0.05	90.10
15570.0	23.00	0.02	0.56	0.02	0.03	90.10
17160.0	21.00	0.01	0.36	0.01	0.02	90.10
18300.0	39.50	0.03	0.94	0.04	0.05	90.10
18360.0	1.00	0.02	0.45	0.02	0.02	90.10
18600.0	5.50	0.02	0.62	0.02	0.03	90.10
18720.0	2.50	0.02	0.57	0.02	0.03	90.10
19740.0	14.00	0.01	0.37	0.01	0.02	90.10
19800.0	1.00	0.02	0.45	0.02	0.02	90.10

STABILITY EXPERIMENT
CHP CELL 6
PRESSURE: 100 psi
PRESSURE CORRECTION FACTOR 1.33

TIME (MIN)	PERM RATE (MLS)	PERM RATE (MLS/MIN)	FLUX RATE (US GFD)	FLUX RATE (M3/M2 D)	PRESS CORR (M3/M2 D)	CUM VOL RED (%)
0						
60.0	115.00	1.92	52.08	2.07	2.75	2.27
220.2	300.00	1.87	50.88	2.02	2.69	8.14
1020.0	444.00	0.56	15.08	0.60	0.80	19.87
1080.0	10.00	0.17	4.53	0.18	0.24	31.09
1140.0	11.00	0.18	4.98	0.20	0.26	32.03
1200.0	12.00	0.20	5.43	0.22	0.29	37.49
1320.0	25.00	0.21	5.66	0.23	0.30	47.52
1380.0	13.00	0.22	5.89	0.23	0.31	52.20
1440.0	13.00	0.22	5.89	0.23	0.31	56.82
1500.0	13.00	0.22	5.89	0.23	0.31	60.27
2460.0	259.00	0.27	7.33	0.29	0.39	67.12
2520.0	18.00	0.30	8.15	0.32	0.43	72.16
2580.0	20.00	0.33	9.06	0.36	0.48	77.26
2640.0	21.00	0.35	9.51	0.38	0.50	82.16
2760.0	23.00	0.19	5.21	0.21	0.28	88.80
2820.0	5.00	0.08	2.26	0.09	0.12	90.10
2940.0	3.50	0.03	0.79	0.03	0.04	90.10
3900.0	22.00	0.02	0.62	0.02	0.03	90.10
4020.0	30.00	0.25	6.79	0.27	0.36	90.10
4200.0	61.50	0.34	9.28	0.37	0.49	90.10
4380.0	51.00	0.28	7.70	0.31	0.41	90.10
5335.2	202.00	0.21	5.75	0.23	0.30	90.10
5788.8	109.00	0.24	6.53	0.26	0.35	90.10
6958.8	242.00	0.21	5.62	0.22	0.30	90.10
8220.0	213.00	0.17	4.59	0.18	0.24	90.10
8340.0	24.50	0.20	5.55	0.22	0.29	90.10
8520.0	36.50	0.20	5.51	0.22	0.29	90.10
8700.0	37.00	0.21	5.58	0.22	0.30	90.10
9660.0	145.00	0.15	4.10	0.16	0.22	90.10
9780.0	22.50	0.19	5.09	0.20	0.27	90.10
9795.0	2.50	0.17	4.53	0.18	0.24	90.10
9900.0	25.00	0.24	6.47	0.26	0.34	90.10
10140.0	52.00	0.22	5.89	0.23	0.31	90.10
11100.0	158.00	0.16	4.47	0.18	0.24	90.10
11220.0	24.00	0.20	5.43	0.22	0.29	90.10
11460.0	50.00	0.21	5.66	0.23	0.30	90.10
11580.0	23.50	0.20	5.32	0.21	0.28	90.10
12540.0	117.00	0.12	3.31	0.13	0.18	90.10
12660.0	20.50	0.17	4.64	0.18	0.25	90.10
12840.0	42.00	0.23	6.34	0.25	0.34	90.10
13020.0	30.50	0.17	4.60	0.18	0.24	90.10
13980.0	120.00	0.13	3.40	0.14	0.18	90.10
14100.0	15.00	0.13	3.40	0.14	0.18	90.10
14280.0	23.50	0.13	3.55	0.14	0.19	90.10
14460.0	22.00	0.12	3.32	0.13	0.18	90.10
15570.0	76.00	0.07	1.86	0.07	0.10	90.10
17160.0	84.00	0.05	1.44	0.06	0.08	90.10
18300.0	179.00	0.16	4.27	0.17	0.23	90.10
18360.0	8.50	0.14	3.85	0.15	0.20	90.10
18600.0	34.50	0.14	3.91	0.16	0.21	90.10
18720.0	18.00	0.15	4.08	0.16	0.22	90.10
19740.0	103.00	0.10	2.74	0.11	0.15	90.10
19800.0	8.00	0.13	3.62	0.14	0.19	90.10

APPENDIX F

RESULTS OF REVERSE OSMOSIS TESTS

EMR INITIAL SALT WATER FLUX TEST
FOR REVERSE OSMOSIS MEMBRANE
SELECTION TEST

TFC CELL1

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00						
60.00	5000.00	710.00	23.67	1.92	78.00	98.44
90.00	3500.00	720.00	24.00	1.95	80.00	97.71
120.00	4000.00	460.00	15.33	1.24	57.00	98.58
150.00	3500.00	375.00	12.50	1.01	44.00	98.74
210.00	4000.00	750.00	12.50	1.01	41.00	98.98
270.00	4000.00	780.00	13.00	1.05	43.00	98.93

CA CELL2

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00						
60.00	5000.00	460.00	15.33	1.24	215.00	95.70
90.00	3500.00	455.00	15.17	1.23	185.00	94.71
120.00	4000.00	290.00	9.67	0.78	155.00	96.13
150.00	3500.00	240.00	8.00	0.65	123.00	96.49
210.00	4000.00	465.00	7.75	0.63	123.00	96.93
270.00	4000.00	500.00	8.33	0.68	123.00	96.93

TFC CELL3

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00						
60.00	5000.00	595.00	19.83	1.61	193.00	96.14
90.00	3500.00	580.00	19.33	1.57	205.00	94.14
120.00	4000.00	375.00	12.50	1.01	152.00	96.20
150.00	3500.00	305.00	10.17	0.83	115.00	96.71
210.00	4000.00	605.00	10.08	0.82	106.00	97.35
270.00	4000.00	640.00	10.67	0.87	113.00	97.18

CA CELL4

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00						
60.00	5000.00	585.00	19.50	1.58	200.00	96.00
90.00	3500.00	470.00	15.67	1.27	185.00	94.71
120.00	4000.00	310.00	10.33	0.84	142.00	96.45
150.00	3500.00	250.00	8.33	0.68	110.00	96.86
210.00	4000.00	480.00	8.00	0.65	103.00	97.43
270.00	4000.00	515.00	8.58	0.70	112.00	97.20

EMR TREATMENT OF UF PERMEATE BY REVERSE OSMOSIS, MEMBRANE SELECTION TEST

TFC CELL1

TIME (MIN)	FEED COND. (umHmos)	FEED pH	PERMEATE pH	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
0.00								
30.00	670.00	8.18	7.36	465.00	15.50	1.26	27.00	95.97
60.00	670.00	8.27	7.15	405.00	13.50	1.10	10.00	98.51
90.00	710.00	8.22	7.55	390.00	13.00	1.05	5.60	99.21
120.00	750.00	8.25	6.85	400.00	13.33	1.08	5.40	99.28
150.00	790.00	8.25	6.75	340.00	11.33	0.92	5.40	99.32
180.00	875.00	8.28	6.68	335.00	11.17	0.91	5.20	99.41
210.00	942.00	8.23	6.75	330.00	11.00	0.89	6.80	99.28
240.00	960.00	8.29	6.65	295.00	9.83	0.80	8.30	99.14
270.00	1040.00	8.50	6.68	305.00	10.17	0.83	9.00	99.13

CA CELL2

TIME (MIN)	FEED COND. (umHmos)	FEED pH	PERMEATE pH	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
0.00								
30.00	670.00	8.18	7.13	310.00	10.33	0.84	47.00	92.99
60.00	670.00	8.27	7.26	275.00	9.17	0.74	14.50	97.84
90.00	710.00	8.22	7.32	260.00	8.67	0.70	11.00	98.45
120.00	750.00	8.25	6.70	255.00	8.50	0.69	10.00	98.67
150.00	790.00	8.25	6.30	230.00	7.67	0.62	11.90	98.49
180.00	875.00	8.28	6.55	220.00	7.33	0.60	10.20	98.83
210.00	942.00	8.23	6.48	215.00	7.17	0.58	11.10	98.82
240.00	960.00	8.29	6.45	185.00	6.17	0.50	13.20	98.63
270.00	1040.00	8.50	6.51	195.00	6.50	0.53	13.60	98.69

TFC CELL3

TIME (MIN)	FEED COND. (umHmos)	FEED pH	PERMEATE pH	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
0.00								
30.00	670.00	8.18	6.95	385.00	12.83	1.04	37.00	94.48
60.00	670.00	8.27	7.45	325.00	10.83	0.88	11.30	98.31
90.00	710.00	8.22	7.18	310.00	10.33	0.84	8.00	98.87
120.00	750.00	8.25	6.65	310.00	10.33	0.84	7.10	99.05
150.00	790.00	8.25	6.45	255.00	8.50	0.69	8.00	98.99
180.00	875.00	8.28	6.46	260.00	8.67	0.70	7.30	99.17
210.00	942.00	8.23	6.50	255.00	8.50	0.69	8.40	99.11
240.00	960.00	8.29	6.55	210.00	7.00	0.57	8.40	99.13
270.00	1040.00	8.50	6.47	225.00	7.50	0.61	10.10	99.03

CA CELL4

TIME (MIN)	FEED COND. (umHmos)	FEED pH	PERMEATE pH	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
0.00								
30.00	670.00	8.18	7.22	340.00	11.33	0.92	57.00	91.49
60.00	670.00	8.27	7.60	300.00	10.00	0.81	15.00	97.76
90.00	710.00	8.22	7.00	280.00	9.33	0.76	11.60	98.37
120.00	750.00	8.25	6.50	285.00	9.50	0.77	11.20	98.51
150.00	790.00	8.25	6.33	235.00	7.83	0.64	12.00	98.48
180.00	875.00	8.28	6.47	235.00	7.83	0.64	12.80	98.54
210.00	942.00	8.23	6.38	230.00	7.67	0.62	14.00	98.51
240.00	960.00	8.29	6.43	200.00	6.67	0.54	14.60	98.48
270.00	1040.00	8.50	6.41	215.00	7.17	0.58	16.50	98.41

CUMMULATIVE VOLUME REDUCTION FOR THE R.O. MEMBRANE SELECTION TEST

TIME (MIN)	TFC CELL1 PERM VOL (MLS)	CA CELL2 PERM VOL (MLS)	TFC CELL3 PERM VOL (MLS)	CA CELL4 PERM VOL (MLS)	TOTAL PERM VOL (MLS)	CUMM. VOL RED (%)
0.00						
30.00	465.00	310.00	385.00	340.00	1500.00	10.71
60.00	405.00	275.00	325.00	300.00	2805.00	20.04
90.00	390.00	260.00	310.00	280.00	4045.00	28.89
120.00	400.00	255.00	310.00	285.00	5295.00	37.82
150.00	340.00	230.00	255.00	235.00	6355.00	45.39
180.00	335.00	220.00	260.00	235.00	7405.00	52.89
210.00	330.00	215.00	255.00	230.00	8435.00	60.25
240.00	295.00	185.00	210.00	200.00	9325.00	66.61
270.00	305.00	195.00	225.00	215.00	10265.00	73.32

EMR POST SALT WATER FLUX TEST
FOR REVERSE OSMOSIS MEMBRANE
SELECTION TEST

TFC CELL1

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
0.00						
30.00	1600.00	385.00	12.83	1.04	98.00	93.88
60.00	3000.00	440.00	14.67	1.19	66.00	97.80
90.00	3500.00	430.00	14.33	1.16	64.00	98.17
120.00	3500.00	430.00	14.33	1.16	51.00	98.54
150.00	3250.00	425.00	14.17	1.15	46.00	98.58

CA CELL2

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
0.00						
30.00	1600.00	250.00	8.33	0.68	118.00	92.63
60.00	3000.00	270.00	9.00	0.73	150.00	95.00
90.00	3500.00	260.00	8.67	0.70	150.00	95.71
120.00	3500.00	250.00	8.33	0.68	145.00	95.86
150.00	3250.00	250.00	8.33	0.68	95.00	97.08

TFC CELL3

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
0.00						
30.00	1600.00	300.00	10.00	0.81	71.00	95.56
60.00	3000.00	350.00	11.67	0.95	86.00	97.13
90.00	3500.00	340.00	11.33	0.92	98.00	97.20
120.00	3500.00	340.00	11.33	0.92	98.00	97.20
150.00	3250.00	340.00	11.33	0.92	96.00	97.05

CA CELL4

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
0.00						
30.00	1600.00	305.00	10.17	0.83	400.00	75.00
60.00	3000.00	285.00	9.50	0.77	223.00	92.57
90.00	3500.00	275.00	9.17	0.74	198.00	94.34
120.00	3500.00	265.00	8.83	0.72	193.00	94.49
150.00	3250.00	260.00	8.67	0.70	180.00	94.46

EMR INITIAL SALT WATER FLUX TEST
FOR REVERSE OSMOSIS OF UF PERMEATE

CELL1 (TFC)

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15.00	5000.00	285.00	19.00	1.54	102.00	97.96
30.00	6700.00	370.00	24.67	2.00	90.00	98.66
45.00	6800.00	455.00	30.33	2.46	98.00	98.56
60.00	5800.00	430.00	28.67	2.33	80.00	98.62
75.00	5500.00	295.00	19.67	1.60	68.00	98.76
90.00	5200.00	275.00	18.33	1.49	60.00	98.85
105.00	5200.00	265.00	17.67	1.43	62.00	98.81
120.00	5200.00	260.00	17.33	1.41	57.00	98.90

CELL2 (TFC)

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15.00	5000.00	255.00	17.00	1.38	190.00	96.20
30.00	6700.00	350.00	23.33	1.89	190.00	97.16
45.00	6800.00	440.00	29.33	2.38	235.00	96.54
60.00	5800.00	385.00	25.67	2.08	205.00	96.47
75.00	5500.00	255.00	17.00	1.38	155.00	97.18
90.00	5200.00	240.00	16.00	1.30	138.00	97.35
105.00	5200.00	230.00	15.33	1.24	130.00	97.50
120.00	5200.00	225.00	15.00	1.22	123.00	97.63

CELL3 (TFC)

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15.00	5000.00	230.00	15.33	1.24	165.00	96.70
30.00	6700.00	300.00	20.00	1.62	158.00	97.64
45.00	6800.00	355.00	23.67	1.92	170.00	97.50
60.00	5800.00	330.00	22.00	1.79	148.00	97.45
75.00	5500.00	220.00	14.67	1.19	113.00	97.95
90.00	5200.00	205.00	13.67	1.11	100.00	98.08
105.00	5200.00	195.00	13.00	1.05	94.00	98.19
120.00	5200.00	195.00	13.00	1.05	91.00	98.25

CELL4 (TFC)

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15.00	5000.00	250.00	16.67	1.35	205.00	95.90
30.00	6700.00	320.00	21.33	1.73	200.00	97.01
45.00	6800.00	430.00	28.67	2.33	205.00	96.99
60.00	5800.00	380.00	25.33	2.06	190.00	96.72
75.00	5500.00	260.00	17.33	1.41	148.00	97.31
90.00	5200.00	230.00	15.33	1.24	130.00	97.50
105.00	5200.00	220.00	14.67	1.19	122.00	97.65
120.00	5200.00	220.00	14.67	1.19	114.00	97.81

CUMMULATIVE VOLUME REDUCTION
RO TREATMENT OF UF PERMEATE

TIME (MIN)	TFC CELL1 PERM VOL (MLS)	TFC CELL2 PERM VOL (MLS)	TFC CELL3 PERM VOL (MLS)	TFC CELL4 PERM VOL (MLS)	TOTAL PERM VOL (MLS)	CUMM. VOL RED (%)
15	250	235	210	235	930	4.13
30	290	270	230	255	1045	8.78
45	300	265	230	265	1060	13.49
60	300	280	240	270	1090	18.33
75	295	270	230	265	1060	23.04
90	285	260	225	255	1025	27.60
105	285	260	225	250	1020	32.13
120	250	225	200	220	895	36.11
135	285	255	220	250	1010	40.60
150	280	250	220	250	1000	45.04
165	275	240	210	240	965	49.33
180	280	240	220	230	970	53.64
195	265	225	200	225	915	57.71
210	275	230	210	235	950	61.93
225	295	235	220	245	995	66.36
240	275	210	200	215	900	70.36
255	325	235	225	260	1045	75.00
270	345	215	210	260	1030	79.58
285	370	195	190	265	1020	84.11
300	360	160	150	240	910	88.16
313	280	90	95	170	635	90.98

EMR TREATMENT BY REVERSE OSMOSIS OF UF PERMEATE

CELL1 (TFC)

TIME (MIN)	FEED COND. (umHmos)	FEED pH	PERMEATE pH	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15	590			250	16.67	1.35	85.0	85.59
30	620	7.85	6.78	290	19.33	1.57	9.8	98.42
45	690			300	20.00	1.62	7.2	98.96
60	700	7.90	6.60	300	20.00	1.62	8.4	98.80
75	750			295	19.67	1.60	8.9	98.81
90	775	7.75	6.43	285	19.00	1.54	6.5	99.16
105	820			285	19.00	1.54	6.0	99.27
120	900	7.85	7.05	250	16.67	1.35	6.2	99.31
135	950			285	19.00	1.54	5.7	99.40
150	975	8.05	6.95	280	18.67	1.51	6.5	99.33
165	970			275	18.33	1.49	7.6	99.22
180	1100	7.95	7.01	280	18.67	1.51	8.3	99.25
195	1210			265	17.67	1.43	7.4	99.39
210	1300	8.13	6.85	275	18.33	1.49	8.4	99.35
225	1450			295	19.67	1.60	8.8	99.39
240	1550	8.20	7.03	275	18.33	1.49	9.9	99.36
255	1900			325	21.67	1.76	11.2	99.41
270	2250	8.48	6.97	345	23.00	1.87	13.0	99.42
285	2750			370	24.67	2.00	15.0	99.45
300	3500	8.60		360	24.00	1.95	19.2	99.45
313	4400			280	21.54	1.75	28.0	99.36

CELL2 (TFC)

TIME (MIN)	FEED COND. (umHmos)	FEED pH	PERMEATE pH	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15	590			235	15.67	1.27	90.0	84.75
30	620	7.85	6.85	270	18.00	1.46	14.8	97.61
45	690			265	17.67	1.43	11.8	98.29
60	700	7.90	6.75	280	18.67	1.51	10.0	98.57
75	750			270	18.00	1.46	9.6	98.72
90	775	7.75	6.40	260	17.33	1.41	10.4	98.66
105	820			260	17.33	1.41	9.3	98.87
120	900	7.85	7.10	225	15.00	1.22	9.6	98.93
135	950			255	17.00	1.38	10.3	98.92
150	975	8.05	7.10	250	16.67	1.35	10.2	98.95
165	970			240	16.00	1.30	11.4	98.82
180	1100	7.95	6.98	240	16.00	1.30	11.1	98.99
195	1210			225	15.00	1.22	15.5	98.72
210	1300	8.13	6.96	230	15.33	1.24	16.0	98.77
225	1450			235	15.67	1.27	16.2	98.88
240	1550	8.20	7.15	210	14.00	1.14	16.5	98.94
255	1900			235	15.67	1.27	18.1	99.05
270	2250	8.48	7.08	215	14.33	1.16	22.0	99.02
285	2750			195	13.00	1.05	24.0	99.13
300	3500	8.60		160	10.67	0.87	34.0	99.03
313	4400			90	6.92	0.56	41.5	99.06

EMR TREATMENT BY REVERSE OSMOSIS OF UF PERMEATE

CELL3 (TFC)

TIME (MIN)	FEED COND. (umHmos)	FEED pH	PERMEATE pH	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15	590			210	14.00	1.14	56.0	90.51
30	620	7.85	6.85	230	15.33	1.24	9.8	98.42
45	690			230	15.33	1.24	8.8	98.72
60	700	7.90	6.85	240	16.00	1.30	9.3	98.67
75	750			230	15.33	1.24	7.1	99.05
90	775	7.75	6.32	225	15.00	1.22	7.6	99.02
105	820			225	15.00	1.22	7.7	99.06
120	900	7.85	7.20	200	13.33	1.08	7.8	99.13
135	950			220	14.67	1.19	10.0	98.95
150	975	8.05	7.05	220	14.67	1.19	8.5	99.13
165	970			210	14.00	1.14	9.1	99.06
180	1100	7.95	7.02	220	14.67	1.19	10.0	99.09
195	1210			200	13.33	1.08	10.7	99.12
210	1300	8.13	7.08	210	14.00	1.14	12.1	99.07
225	1450			220	14.67	1.19	12.8	99.12
240	1550	8.20	7.16	200	13.33	1.08	13.8	99.11
255	1900			225	15.00	1.22	17.3	99.09
270	2250	8.48	7.05	210	14.00	1.14	21.5	99.04
285	2750			190	12.67	1.03	26.2	99.05
300	3500	8.60		150	10.00	0.81	28.5	99.19
313	4400			95	7.31	0.59	43.5	99.01

CELL4 (TFC)

TIME (MIN)	FEED COND. (umHmos)	FEED pH	PERMEATE pH	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15	590			235	15.67	1.27	129.0	78.14
30	620	7.85	7.03	255	17.00	1.38	14.6	97.65
45	690			265	17.67	1.43	14.2	97.94
60	700	7.90	7.01	270	18.00	1.46	17.3	97.53
75	750			265	17.67	1.43	8.8	98.83
90	775	7.75	6.33	255	17.00	1.38	8.7	98.88
105	820			250	16.67	1.35	9.8	98.80
120	900	7.85	7.20	220	14.67	1.19	9.2	98.98
135	950			250	16.67	1.35	9.9	98.96
150	975	8.05	7.17	250	16.67	1.35	9.8	98.99
165	970			240	16.00	1.30	11.0	98.87
180	1100	7.95	7.20	230	15.33	1.24	12.2	98.89
195	1210			225	15.00	1.22	12.4	98.98
210	1300	8.13	7.10	235	15.67	1.27	15.5	98.81
225	1450			245	16.33	1.33	16.0	98.90
240	1550	8.20	7.23	215	14.33	1.16	15.0	99.03
255	1900			260	17.33	1.41	18.0	99.05
270	2250	8.48	7.15	260	17.33	1.41	17.0	99.24
285	2750			265	17.67	1.43	23.0	99.16
300	3500	8.60		240	16.00	1.30	24.8	99.29
313	4400			170	13.08	1.06	38.0	99.14

EMR POST SALT WATER FLUX TEST
FOR REVERSE OSMOSIS OF UF PERMEATE

CELL1 (TFC)

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15.00	4700.00	220.00	14.67	1.19	130.00	97.23
30.00	4800.00	250.00	16.67	1.35	68.00	98.58
45.00	4900.00	260.00	17.33	1.41	54.00	98.90
60.00	4900.00	250.00	16.67	1.35	52.50	98.93
75.00	4950.00	250.00	16.67	1.35	51.50	98.96
90.00	4830.00	245.00	16.33	1.33	50.00	98.96
105.00	4900.00	245.00	16.33	1.33	50.00	98.98
120.00	4820.00	245.00	16.33	1.33	50.00	98.96

CELL2 (TFC)

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15.00	4700.00	165.00	11.00	0.89	165.00	96.49
30.00	4800.00	126.00	8.40	0.68	126.00	97.38
45.00	4900.00	125.00	8.33	0.68	125.00	97.45
60.00	4900.00	127.00	8.47	0.69	127.00	97.41
75.00	4950.00	129.00	8.60	0.70	129.00	97.39
90.00	4830.00	123.00	8.20	0.67	123.00	97.45
105.00	4900.00	120.00	8.00	0.65	120.00	97.55
120.00	4820.00	125.00	8.33	0.68	125.00	97.41

CELL3 (TFC)

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15.00	4700.00	100.00	6.67	0.54	268.00	94.30
30.00	4800.00	110.00	7.33	0.60	160.00	96.67
45.00	4900.00	110.00	7.33	0.60	130.00	97.35
60.00	4900.00	110.00	7.33	0.60	121.00	97.53
75.00	4950.00	105.00	7.00	0.57	110.00	97.78
90.00	4830.00	105.00	7.00	0.57	111.00	97.70
105.00	4900.00	105.00	7.00	0.57	105.00	97.86
120.00	4820.00	105.00	7.00	0.57	104.00	97.84

CELL4 (TFC)

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
15.00	4700.00	155.00	10.33	0.84	285.00	93.94
30.00	4800.00	165.00	11.00	0.89	185.00	96.15
45.00	4900.00	170.00	11.33	0.92	150.00	96.94
60.00	4900.00	160.00	10.67	0.87	147.00	97.00
75.00	4950.00	160.00	10.67	0.87	146.00	97.05
90.00	4830.00	155.00	10.33	0.84	145.00	97.00
105.00	4900.00	160.00	10.67	0.87	140.00	97.14
120.00	4820.00	160.00	10.67	0.87	130.00	97.30

EMR INITIAL SALT WATER FLUX TEST
FOR REVERSE OSMOSIS TEST ON
DEMULSIFIED PRODUCED WATER

CELL1 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00	1560.00	360.00	12.00	0.97	53.00	96.60
60.00	2000.00	365.00	12.17	0.99	44.00	97.80
90.00	3000.00	360.00	12.00	0.97	40.00	98.67
120.00	3100.00	360.00	12.00	0.97	38.00	98.77

CELL2 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00	1560.00	370.00	12.33	1.00	61.00	96.09
60.00	2000.00	360.00	12.00	0.97	52.00	97.40
90.00	3000.00	360.00	12.00	0.97	49.00	98.37
120.00	3100.00	355.00	11.83	0.96	46.00	98.52

CELL3 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00	1560.00	360.00	12.00	0.97	47.00	96.99
60.00	2000.00	360.00	12.00	0.97	39.00	98.05
90.00	3000.00	345.00	11.50	0.93	36.00	98.80
120.00	3100.00	345.00	11.50	0.93	33.00	98.94

CELL4 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00	1560.00	355.00	11.83	0.96	49.00	96.86
60.00	2000.00	355.00	11.83	0.96	40.00	98.00
90.00	3000.00	345.00	11.50	0.93	39.00	98.70
120.00	3100.00	340.00	11.33	0.92	34.00	98.90

CUMULATIVE VOLUME REDUCTION
RO TREATMENT OF DEMULSIFIED PRODUCED WATER

TIME (MIN)	TFC CELL1 PERM VOL (MLS)	TFC CELL2 PERM VOL (MLS)	TFC CELL3 PERM VOL (MLS)	TFC CELL4 PERM VOL (MLS)	TOTAL PERM VOL (MLS)	CUMM. VOL RED (%)
30	285	295	280	285	1145	5.73
60	305	300	295	290	1190	11.68
90	300	290	280	285	1155	17.45
120	295	280	275	280	1130	23.10
150	295	280	270	270	1115	28.68
180	290	265	260	265	1080	34.08
210	290	265	260	260	1075	39.45
240	285	255	250	255	1045	44.68
270	295	245	245	250	1035	49.85
300	280	245	240	245	1010	54.90
330	290	250	250	245	1035	60.07
360	310	255	250	255	1070	65.42
390	310	260	255	255	1080	70.82
420	310	250	250	275	1085	76.25
435	160	125	130	135	550	79.00

EMR TREATMENT OF UF PERMEATE
BY REVERSE OSMOSIS ON
DEMULSIFIED PRODUCED WATER

CELL3 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30	1100	280	9.33	0.76	26.5	97.59
60	1150	295	9.83	0.80	10.4	99.10
90	1200	280	9.33	0.76	11.8	99.02
120	1250	275	9.17	0.74	10.1	99.19
150	1400	270	9.00	0.73	11.6	99.17
180	1500	260	8.67	0.70	11.1	99.26
210	1630	260	8.67	0.70	11.9	99.27
240	1800	250	8.33	0.68	12.9	99.28
270	1900	245	8.17	0.66	14.0	99.26
300	2250	240	8.00	0.65	17.5	99.22
330	1950	250	8.33	0.68	17.5	99.10
360	1500	250	8.33	0.68	23.0	98.47
390	3500	255	8.50	0.69	26.0	99.26
420	5000	250	8.33	0.68	38.0	99.24
435	4950	130	8.67	0.70	30.0	99.39

CELL4 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30	1100	285	9.50	0.77	26.5	97.59
60	1150	290	9.67	0.78	10.4	99.10
90	1200	285	9.50	0.77	11.8	99.02
120	1250	280	9.33	0.76	10.1	99.19
150	1400	270	9.00	0.73	11.6	99.17
180	1500	265	8.83	0.72	11.1	99.26
210	1630	260	8.67	0.70	11.9	99.27
240	1800	255	8.50	0.69	12.9	99.28
270	1900	250	8.33	0.68	14.0	99.26
300	2250	245	8.17	0.66	17.5	99.22
330	1950	245	8.17	0.66	17.5	99.10
360	1500	255	8.50	0.69	23.0	98.47
390	3500	255	8.50	0.69	26.0	99.26
420	5000	275	9.17	0.74	38.0	99.24
435	4950	135	9.00	0.73	30.0	99.39

EMR TREATMENT OF UF PERMEATE BY REVERSE OSMOSIS ON DEMULSIFIED PRODUCED WATER

CELL1 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30	1100	285	9.50	0.77	26.5	97.59
60	1150	305	10.17	0.83	10.4	99.10
90	1200	300	10.00	0.81	11.8	99.02
120	1250	295	9.83	0.80	10.1	99.19
150	1400	295	9.83	0.80	11.6	99.17
180	1500	290	9.67	0.78	11.1	99.26
210	1630	290	9.67	0.78	11.9	99.27
240	1800	285	9.50	0.77	12.9	99.28
270	1900	295	9.83	0.80	14.0	99.26
300	2250	280	9.33	0.76	17.5	99.22
330	1950	290	9.67	0.78	17.5	99.10
360	1500	310	10.33	0.84	23.0	98.47
390	3500	310	10.33	0.84	26.0	99.26
420	5000	310	10.33	0.84	38.0	99.24
435	4950	160	10.67	0.87	30.0	99.39

CELL2 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30	1100	295	9.83	0.80	26.5	97.59
60	1150	300	10.00	0.81	10.4	99.10
90	1200	290	9.67	0.78	11.8	99.02
120	1250	280	9.33	0.76	10.1	99.19
150	1400	280	9.33	0.76	11.6	99.17
180	1500	265	8.83	0.72	11.1	99.26
210	1630	265	8.83	0.72	11.9	99.27
240	1800	255	8.50	0.69	12.9	99.28
270	1900	245	8.17	0.66	14.0	99.26
300	2250	245	8.17	0.66	17.5	99.22
330	1950	250	8.33	0.68	17.5	99.10
360	1500	255	8.50	0.69	23.0	98.47
390	3500	260	8.67	0.70	26.0	99.26
420	5000	250	8.33	0.68	38.0	99.24
435	4950	125	8.33	0.68	30.0	99.39

EMR POST SALT WATER FLUX TEST FOR REVERSE OSMOSIS TEST ON DEMULSIFIED PRODUCED WATER

CELL1 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00	2900.00	405.00	13.50	1.10	69.00	97.62
60.00	2760.00	405.00	13.50	1.10	52.00	98.12
90.00	3280.00	450.00	15.00	1.22	53.00	98.38
120.00	2820.00	445.00	14.83	1.20	60.00	97.87

CELL2 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00	2900.00	380.00	12.67	1.03	69.00	97.62
60.00	2760.00	375.00	12.50	1.01	65.00	97.64
90.00	3280.00	410.00	13.67	1.11	68.00	97.93
120.00	2820.00	405.00	13.50	1.10	70.20	97.51

CELL3 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00	2900.00	360.00	12.00	0.97	42.10	98.55
60.00	2760.00	355.00	11.83	0.96	42.40	98.46
90.00	3280.00	385.00	12.83	1.04	41.50	98.73
120.00	2820.00	390.00	13.00	1.05	49.90	98.23

CELL4 TFC

TIME (MIN)	FEED COND. (umHmos)	PERMEATE RATE (MLS)	PERMEATE RATE (MLS/MIN)	FLUX RATE (M3/M2 D)	PERMEATE COND. (umHmos)	PERCENT REJECTION (%)
30.00	2900.00	375.00	12.50	1.01	72.50	97.50
60.00	2760.00	375.00	12.50	1.01	50.10	98.18
90.00	3280.00	400.00	13.33	1.08	49.00	98.51
120.00	2820.00	410.00	13.67	1.11	52.00	98.16

APPENDIX G

COSTS OF COMPONENTS OF PRODUCED WATER CLEANING PROCESSES

G.1 **Pretreatment for Ultrafiltration Units**

G.1.1 Design Parameters

Skim tank with surface oil removal, periodic sludge take off and gas blanket.

Flowrate = 540 US.GPM

Nominal Retention Time = 2 hours

G.1.2 Equipment List and Purchase Costs

Skim tank 64,800 US.GAL.	$176,500
Purchase cost of equipment	$176,500

G.1.3 Total Cost of Installed Equipment

Purchased cost of equipment		$ 176,500	
Shipping charges		17,650	
Installation of equipment		14,120	
Foundations		35,300	
Instruments		8,825	
Piping		35,300	
Insulation		35,300	
Electrical		8,825	
Buildings & structures		5,295	
Services		14,120	
Misc.		8,825	
Taxes	Federal	36,006	
TOTAL DIRECT COSTS			396,066
Construction and field expenses		39,607	
Engineering expenses		29,705	
Contractors fees		23,269	
Contingencies		46,538	
TOTAL INDIRECT COSTS			$ 139,118
TOTAL COSTS			$ 535,184

G.1.4 Operating and Maintenance Costs

Miscellaneous	$1,000/year
Contingencies	$1,000/year
Total operating and maintenance costs	$2,000/year

G.2 **Spiral UF System 1**

(not followed by RO)

G.2.1 Design Parameters

Feed flowrate = 625,000 US.GPD = 434 GPM

Permeate flowrate = 500,000 US.GPD = 347 GPM

Concentrate flowrate = 125,000 US.GPD = 87 GPM

Percent recovery = 80%

Operating temp. = 65°C

Operating pressure = 100 psi

Membrane type = 8" spiral

Membrane surface area = 300 ft^2/membrane

Assumed flux rate = $0.7 \times 0.55\ M^3/M^2 \cdot D$

$= 0.385\ M^3/M^2D$

$= 9.45\ USG/F^2 \cdot D$

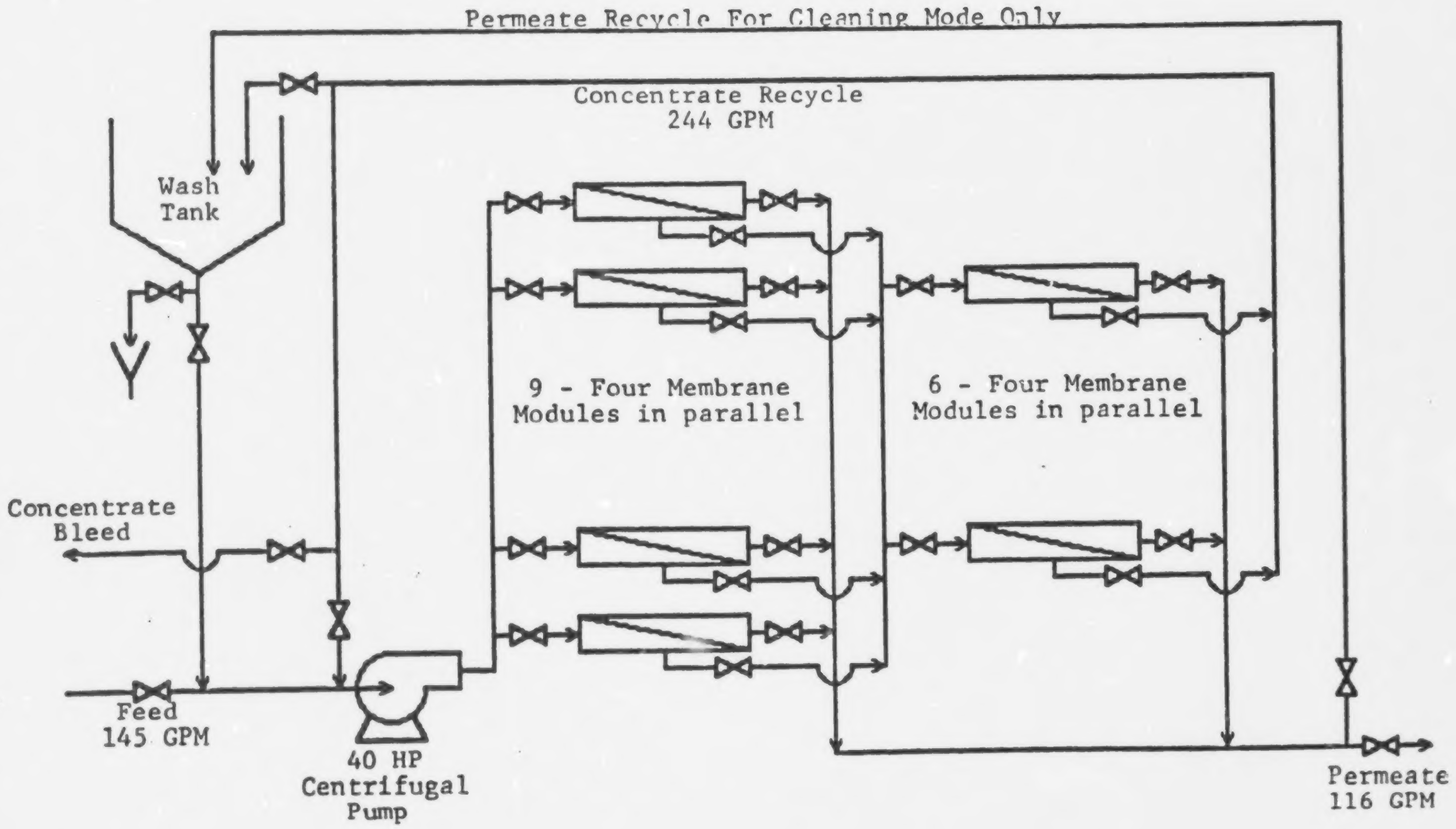
Number of membranes required = 180

Three units each with 15 - 4 membrane modules are required.

G.2.2 Equipment List and Purchase Cost

Purchase cost of each of three, 60 membrane, systems (engineering and assembly work completed)	$291,319
Purchase cost of equipment	$873,957

FIGURE G-1 SPIRAL ULTRAFILTRATION FOR PROCESS FOLLOWED BY SODIUM ZEOLITE SOFTENING. (One of three units required.)



Permeate Recycle For Cleaning Mode Only
Concentrate Recycle
244 GPM
Wash
Tank
9 - Four Membrane
Modules in parallel
6 - Four Membrane
Modules in parallel
Concentrate
Bleed
Feed
145 GPM
40 HP
Centrifugal
Pump
Permeate
116 GPM

G.2.3 Total Cost of Installed Equipment

Purchased cost of equipment	$ 873,957	
Shipping charges	43,698	
Installation of equipment	53,437	
Foundations	43,698	
Instruments	52,437	
Piping	43,698	
Insulation	87,396	
Electrical	69,917	
Buildings & Structures	87,396	
Services	69,917	
Misc.	43,698	
Taxes Federal		146,825
TOTAL DIRECT COSTS	$1,615,073	
Construction and field expenses	80,754	
Engineering expenses	96,904	
Contractors fees	89,637	
Contingencies	179,273	
TOTAL INDIRECT COSTS		$ 446,568
TOTAL COSTS		$2,061,640

G.2.4 Operating and Maintenance Costs

Materials	
Cleaning chemicals (wash every 3 days)	$51,330/year
Replacement membranes (every 3 years)	$98,658/year
Labour (shared with other parts of plant) (1/2 shift/day at $135/day)	$16,875/year
Energy	
Electrical Power (120 HP x 0.7457 = 89.5 KW x 24 x 365 x 1.35 cents/KW)	$10,580/year
Miscellaneous	$5,000/year
Contingency	$18,000/year
Total Operating Costs	$200,443/year

G.3 **Tubular UF System 1**

(not followed by RO)

G.3.1 Design Parameters

Feed rate = 625,000 US.GPD = 434 GPM

Permeate flowrate = 500,000 US.GPD = 347 GPM

Concentrate flowrate = 125,000 US.GPD = 87 GPM

Percent recovery = 80%

Operating temperature = 65°C

Operating pressure = 100 psi

Membrane type = 1/2 inch tubular Nito P18 modules

Membrane surface area = 19.4 ft^2/module

Assumed flux rate = 0.55 $M^3/M^2 \cdot D$ = 13.5 $USG/F^2 \cdot D$

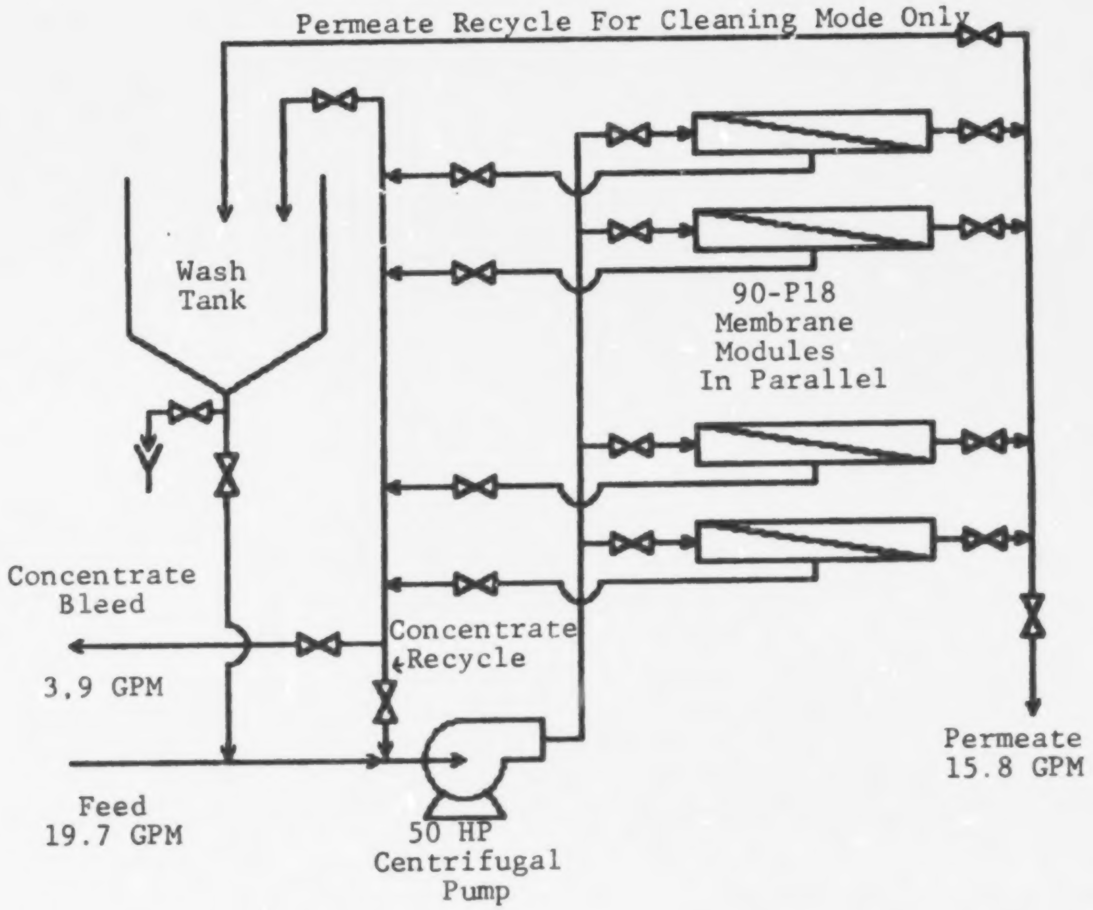
Number of modules required = 1909

22 units each with 90 modules are required.

G.3.2 Equipment List and Costs

Purchase cost of each of 22 membrane systems (engineering and assembly work completed)	173,340
Purchase cost of equipment	$3,813,480

FIGURE G-2 TUBULAR ULTRAFILTRATION SYSTEM FOR PROCESS FOLLOWED BY SODIUM ZEOLITE SOFTENING. (One of 22 units required)


Permeate Recycle For Cleaning Mode Only
Wash Tank
90-P18 Membrane Modules In Parallel
Concentrate Bleed
3.9 GPM
Concentrate Recycle
Permeate 15.8 GPM
Feed 19.7 GPM
50 HP Centrifugal Pump

G.3.3 Total Cost of Installed Equipment

Purchased cost of equipment		$ 3,813,480	
Shipping charges		190,674	
Installation of equipment		190,674	
Foundations		190,674	
Instruments		190,674	
Piping		190,674	
Insulation		381,348	
Electrical		305,078	
Buildings & structures		381,348	
Serivces		305,078	
Misc.		190,674	
Taxes	Federal	633,038	
TOTAL DIRECT COSTS			$6,963,414
Construction and field expenses		348,171	
Engineering expenses		417,805	
Contractors fees		386,470	
Contingencies		772,939	
TOTAL INDIRECT COSTS			$1,925,384
TOTAL COSTS			$8,888,799

G.3.4 Operating and Maintenance Costs

Materials	
Cleaning chemicals (wash every 4 days)	$225,844/year
Replacement membranes (every 3 years)	$297,000/year
Labour (shared with other parts of plant)	$33,750/year
(1 shift/day at $135/day)	
Energy	
Electrical Power	$97,000/year
(110 HP x 0.7457 x 24 x 365 x 1.35 cents/KW)	
Miscellaneous	$6,000/year
Contingency	$33,000/year
Total Operating Costs	$692,594/year

G.4 Post UF Softening

G.4.1 Design Parameters

Flow rate = 350 US.GPM

Feed hardness = 2.5 ppm

Silica concentration = 65 ppm

Total dissolved solids = 1600 ppm

Effluent hardness = 0.5 ppm

System consists of two 5.5 foot diameter sodium zeolite softeners

G.4.2 Equipment List and Purchase Costs

2 - 5.5 foot diameter sodium zeolite softener units = $60,000

Purchase cost of equipment = $60,000

G.4.3 Total Cost of Installed Equipment

Purchased cost of equipment		$ 60,000	
Shipping charges		6,000	
Installation of equipment		9,000	
Foundations		6,000	
Instruments		6,000	
Piping		18,000	
Insulation		12,000	
Electrical		4,800	
Buildings & structures		6,000	
Serivces		7,200	
Misc.		3,000	
Taxes	Federal	13,800	
TOTAL DIRECT COSTS			151,800
Construction and field expenses		15,180	
Engineering expenses		11,385	
Contractors fees		8,918	
Contingencies		17,837	
TOTAL INDIRECT COSTS			53,320
TOTAL COSTS			$205,120

G.4.4 Operating and Maintenance Costs

Materials	
Salt for regeneration	$4,300/year
Labour (shared with other parts of plant)	$16,875/year
(1/2 shift/day at $135/day)	
Miscellaneous	$2,000/year
Contingency	$2,300/year
Total operating costs	$25,475/year

G.5 Post Treatment Storage and O_2 Removal

G.5.1 Design Parameters

Storage tank with a gas blanket and addition of sodium sulfite as an oxygen scavenger.

Flowrate = 350 US.GPM

Storage tank volume = 500,000 US.GAL

Oxygen scavenger concentration added = 50 ppm

G.5.2 Equipment List and Purchase Costs

Storage tank (500,000 US.GAL)	$71,000
Oxygen scavenger addition equipment	$5,000
Purchase cost of equipment	$76,000

G.5.3 Total Cost of Installed Equipment

Purchased cost of equipment		$ 76,000	
Shipping charges		7,600	
Installation of equipment		11,400	
Foundations		15,200	
Instruments		6,080	
Piping		15,200	
Insulation		15,200	
Electrical		5,320	
Buildings & structures		6,080	
Serivces		6,080	
Misc.		3,800	
Taxes	Federal	16,796	
TOTAL DIRECT COSTS			$184,756
Construction and field expenses		18,476	
Engineering expenses		13,857	
Contractors fees		10,854	
Contingencies		21,709	
TOTAL INDIRECT COSTS			$ 64,896
TOTAL COSTS			$249,652

G.5.4 Operating and Maintenance Costs

Materials	
Sodium sulfite (Corrogen)	
50 ppm @ $1.80/Kg	$62,000/year
Labour (1/2 shift/day at $135/day)	$16,875/year
Miscellaneous	$ 7,400/year
Contingencies	$ 8,000/year
Total operating and maintenance cost	$94,275/year

G.6 Spiral UF System 2

(followed by RO)

G.6.1 Design Parameters

Feed flowrate = 781,250 US.GPD = 542.5 US.GPM

Permeate flowrate = 625,000 US.GPD = 434 US.GPM

Concentrate flowrate = 156,250 US.GPD = 108.5 US.GPM

Percent recovery = 80%

Operating temperature = 65°C

Operating pressure = 100 psi

Membrane type = 8" spiral

Membrane surface area = 300 ft^2/membrane

Assumed flux rate = 0.7 x 0.55 $M^3/M^2 \cdot D$

= 0.385 $M^3/M^2 \cdot D$

= 9.45 $USG/F^2 \cdot D$

Number of membranes required = 224

Number of membranes per module = 4

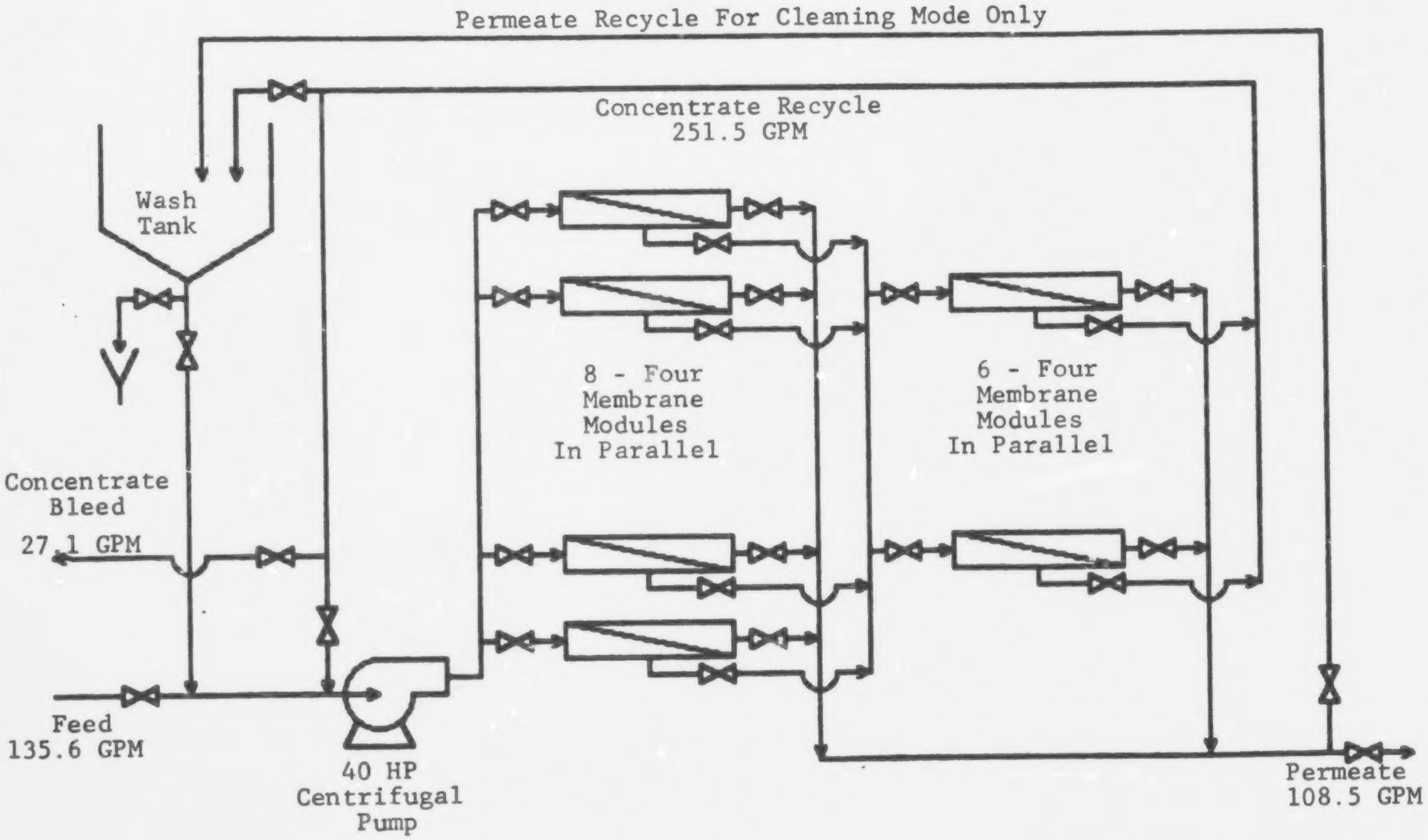
Number of modules = 56

Four units each with 14 - 4 membrane modules are required.

G.6.2 Equipment List and Purchase Costs

Purchase cost of each of four, 60 membrane, systems (engineering and assembly work completed)	$276,921
Purchase cost of equipment	$1,107,684

FIGURE G-3 SPIRAL ULTRAFILTRATION SYSTEM FOR PROCESS FOLLOWED BY REVERSE OSMOSIS.
(One of four units required)


Permeate Recycle For Cleaning Mode Only
Concentrate Recycle
251.5 GPM
Wash
Tank
8 - Four
Membrane
Modules
In Parallel
6 - Four
Membrane
Modules
In Parallel
Concentrate
Bleed
27.1 GPM
Feed
135.6 GPM
40 HP
Centrifugal
Pump
Permeate
108.5 GPM

G.6.3 Total Cost of Installed Equipment

Purchased cost of equipment		$ 1,107,684	
Shipping charges		55,384	
Installation of equipment		66,461	
Foundations		55,384	
Instruments		66,461	
Piping		55,384	
Insulation		110,768	
Electrical		88,615	
Buildings & structures		110,768	
Serivces		88,615	
Misc.		55,384	
Taxes	Federal	186,091	
TOTAL DIRECT COSTS			$2,047,000
Construction and field expenses		102,350	
Engineering expenses		122,820	
Contractors fees		113,609	
Contingencies		227,217	
TOTAL INDIRECT COSTS			$ 565,996
TOTAL COSTS			$2,612,996

G.6.4 Operating and Maintenance Costs

Materials	
Cleaning Chemicals (wash every 3 days)	$68,440/year
Replacement Membranes (every 3 years)	$122,773/year
Labour (Shared with other parts of plant)	$16,875/year
(1/2 shift/day at $135/day)	$16,875/year
Energy	
Electrical Power	$14,110/year
(160 HP x 0.7457 x 24 x 365 x 1.35 cents/KW)	
Miscellaneous	$4,000/year
Contingency	$22,500/year
Total Operating Costs	$248,698/year

G.7 **Tubular UF System**

(followed by RO)

G.7.1 Design Parameters

Feed flowrate = 781,250 US.GPD = 542.5 US.GPM

Permeate flowrate = 625,000 US.GPD = 434 US.GPM

Concentrate flowrate = 156,250 US.GPD = 108.5 US.GPM

Percent recovery = 80%

Operating temperature = 65°C

Operating pressure = 100 psi

Membrane type = 1/2" tubular Nito P18 modules

Membrane surface area = 19.4 ft^2/module

Assumed flux rate = 0.55 $M^3/M^2 \cdot D$

= 13.5 $USG/F^2 \cdot D$

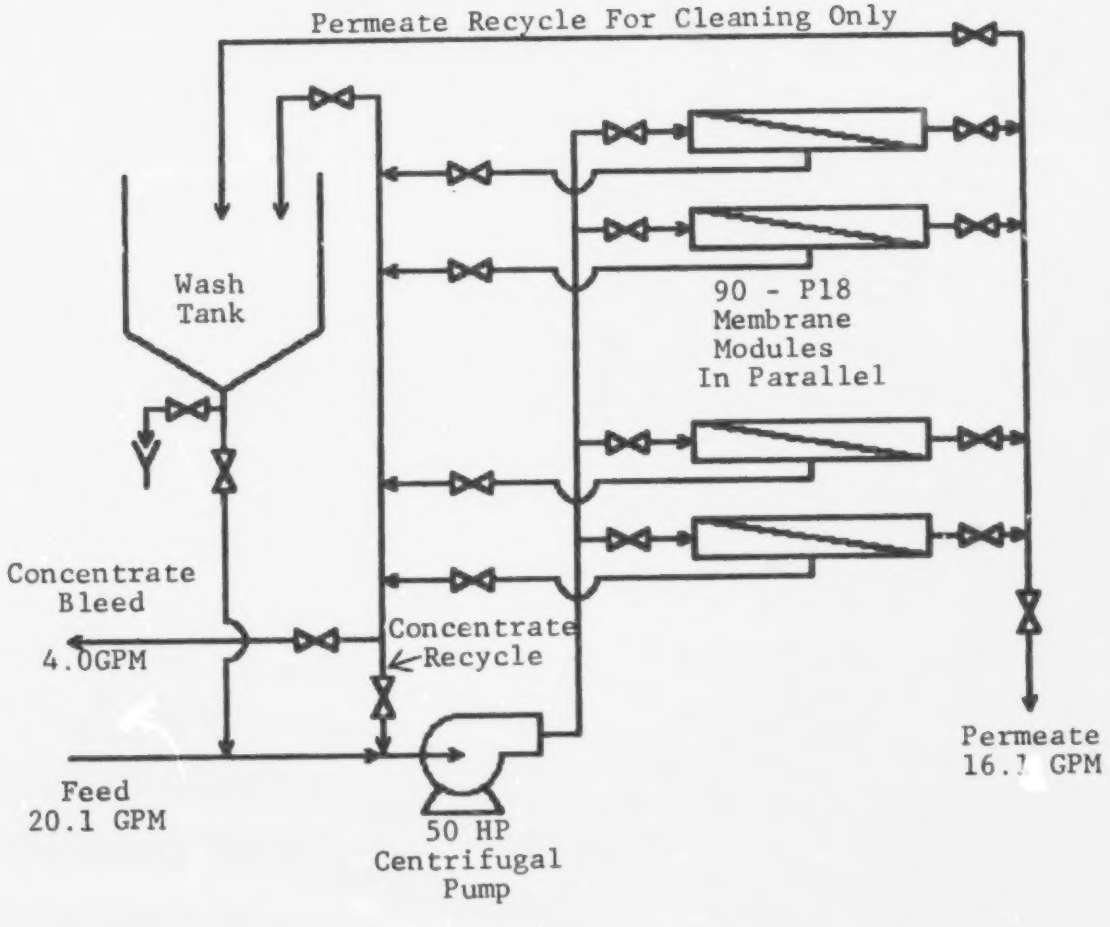
Number of membranes required = 2386

27 units, each with 90 modules, are required.

G.7.2 Equipment List and Purchase Costs

Purchase cost of each of 27 membrane systems (engineering and assembly work completed)	$173,340
Purchase cost of equipment	$4,680,180

FIGURE G-4 TUBULAR ULTRAFILTRATION SYSTEM FOR PROCESS FOLLOWED BY REVERSE OSMOSIS.
(One of 27 units required)


Permeate Recycle For Cleaning Only
Wash Tank
90 - P18 Membrane Modules In Parallel
Concentrate Bleed
4.0GPM
Concentrate Recycle
Permeate
16.1 GPM
Feed
20.1 GPM
50 HP Centrifugal Pump

G.7.3 Total Cost of Installed Equipment

Purchased cost of equipment		$ 4,680,180	
Shipping charges		234,009	
Installation of equipment		280,811	
Foundations		234,009	
Instruments		280,811	
Piping		234,009	
Insulation		468,018	
Electrical		374,414	
Buildings & structures		468,018	
Serivces		374,414	
Misc.		234,009	
Taxes	Federal	$786,270	
TOTAL DIRECT COSTS			$8,648,973
Construction and field expenses		432,449	
Engineering expenses		518,938	
Contractors fees		480,018	
Contingencies		960,036	
TOTAL INDIRECT COSTS			$2,391,441
TOTAL COSTS			$11,040,414

G.7.4 Operating and Maintenance Costs

Materials	
Cleaning Chemicals (wash every 3 days)	$277,172/year
Replacement Membranes (every 3 years)	$364,500/year
Labour (shared with other parts of plant)	
(shift/day at $135/day)	$33,750/year
Energy	
Electrical Power	$119,050/year
(1350 HP x 0.7457 x 24 x 365 x 1.35 cents/KW)	
Miscellaneous	$6,000/year
Contingency	$80,000/year
Total Operating Costs	$880,472/year

G.8 **Spiral RO System for UF Permeate**

G.8.1 Design Parameters

Feed flowrate = 625,000 US.GPD = 434 US.GPM

Permeate flowrate = 500,000US.GPD = 347 US.GPM

Concentrate flowrate = 125,000 US.GPD = 87 US.GPM

Percent recovery = 80%

Operating temperature = 35°C

Operating pressure = 600 psi

Membrane type = 8" diameter x 40" spiral TFC Seawater

Membrane surface area = 350 ft^2/module

Assumed flux rate = $1.0 M^3/M^2 \cdot D$

= $24.6\ USG/F^2 \cdot D$

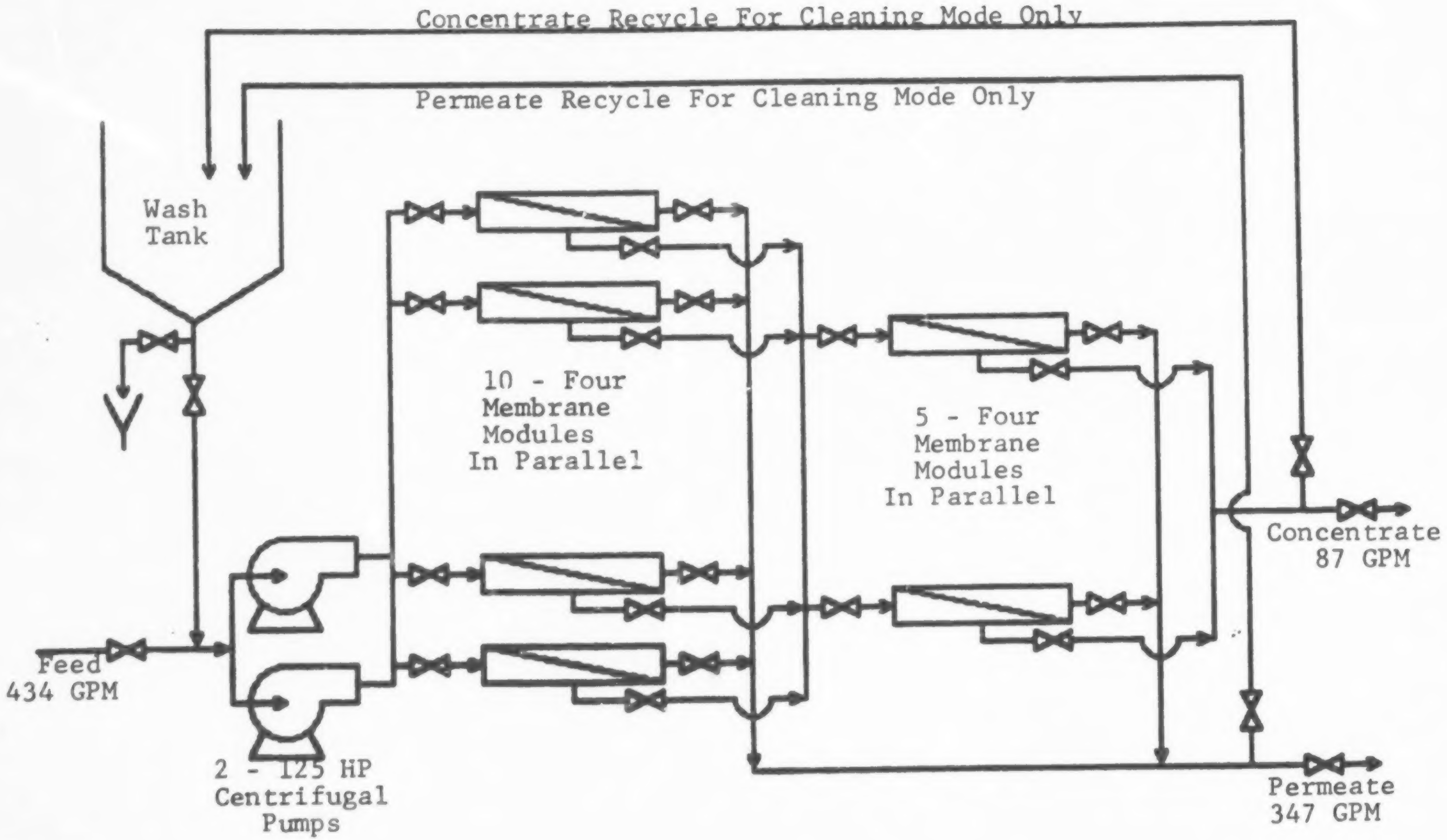
Number of membranes required = 58

15 four membrane modules will be required.

G.8.2 Equipment List and Purchase Costs

Purchase cost of one 58 membrane system	
(engineering and assembly work completed)	$429,859
Purchase cost of equipment	$429,859

FIGURE G-5 SPIRAL REVERSE OSMOSIS SYSTEM FOR ULTRAFILTRATION PERMEATE.


Concentrate Recycle For Cleaning Mode Only
Permeate Recycle For Cleaning Mode Only
Wash Tank
10 - Four Membrane Modules In Parallel
5 - Four Membrane Modules In Parallel
Concentrate 87 GPM
Feed 434 GPM
2 - 125 HP Centrifugal Pumps
Permeate 347 GPM

G.8.3 Total Cost of Installed Equipment

Purchased cost of equipment		$ 429,859	
Shipping charges		21,493	
Installation of equipment		25,792	
Foundations		21,493	
Instruments		25,792	
Piping		21,493	
Insulation		30,090	
Electrical		34,389	
Buildings & structures		42,986	
Serivces		34,389	
Misc.		21,493	
Taxes	Federal	$70,927	
TOTAL DIRECT COSTS			$780,194
Construction and field expenses		39,010	
Engineering expenses		46,812	
Contractors fees		43,301	
Contingencies		86,602	
TOTAL INDIRECT COSTS			$215,724
TOTAL COSTS			$995,918

G.8.4 Operating and Maintenance Costs

Materials	
Cleaning Chemicals (wash every 2 days)	$25,550/year
Replacement Membranes (every 3 years)	$36,540/year
Labour (shared with other parts of plant)	
(1/2 shift/day at $135/day)	$16,875/year
Energy	
Electrical Power	$22,050/year
(250 HP x 0.7457 x 24 x 365 x 1.35 cents/KW)	
Miscellaneous	$4,000/year
Contingency	$10,500/year
Total Operating Costs	$115,515/year

G.9 Pretreatment for Reverse Osmosis Unit

G.9.1 Design Parameters

System consists of; 1) skim tank with demulsifier addition, oil skim, and sludge take off. 2) induced gas flotation (IGF), 3) intermediate storage tank with a gas blanket, 4) sand filter, and 5) cartridge filter unit.

Feed flowrate = 434 US.GPD

Demulsified = RTC 323 (200 ppm)

Skim tank nominal retention time = 2 h

Skim tank effluent oil concentration = 20 - 30 ppm

Storage tank volume = 125,000 US.GAL

Sand filter effluent oil concentration = 5 - 8 ppm

Cartridge filter effluent oil concentration = 1 - 2 ppm

G.9.2 Equipment List and Purchase Costs

Demulsifier addition equipment	$5,000
Skim tank 46,000 US gal.	$180,000
Induced gas flotation unit, Wimco#66	$130,000
Intermediate storage tank	$45,500
Sand filter, Eimco 2 8' diameter units	$60,000
Cartridge filter, Wix facet filter 34R3-CF30CIH	$8,000
Purchase cost of equipment	$428,500

G.9.3 Total Cost of Installed Equipment

Purchased cost of equipment		$ 428,500	
Shipping charges		42,850	
Installation of equipment		64,275	
Foundations		42,850	
Instruments		42,850	
Piping		128,550	
Insulation		85,700	
Electrical		34,280	
Buildings & structures		42,850	
Serivces		51,420	
Misc.		21,425	
Taxes	Federal	$98,555	
TOTAL DIRECT COSTS			$1,084,105
Construction and field expenses		$108,411	
Engineering expenses		$81,308	
Contractors fees		$63,691	
Contingencies		$127,382	
TOTAL INDIRECT COSTS			$ 380,792
TOTAL COSTS			$1,464,897

G.9.4 Operating and Maintenance Costs

Materials	
Demulsifier RTC 323, 200 ppm @ $2.00/Kg	$345,381/year
Cartridge filter replacement	
(34 filters, 12 times/yr, @ $9.30 each)	$3,794/year
Labour (4 shifts/day at $135/day)	$135,000/year
Energy	
Electrical Power	
(skim tank 10 HP, IGF=30 HP, pump at 40HP)	
(80 HP x 0.7457 x 24 x 365 x 1.35 cents/KW)	$7,055/year
Miscellaneous	$10,000/year
Contingency	$50,000/year
Total Operating Costs	$551,230/year

G.10 Spiral RO System for IGF and Filtered Water

G.10.1 Design Parameters

Feed flowrate = 625,000 US.GPD = 434 US.GPM

Permeate flowrate = 500,000US.GPD = 347 US.GPM

Concentrate flowrate = 125,000 US.GPD = 87 US.GPM

Percent recovery = 80%

Operating temperature = 35°C

Operating pressure = 600 psi

Membrane type = 8" diameter x 40" spiral TFC Seawater

Membrane surface area = 350 ft^2/module

Assumed flux rate = 0.775 $M^3/M^2 \cdot D$

= 19.8 $USG/F^2 \cdot D$

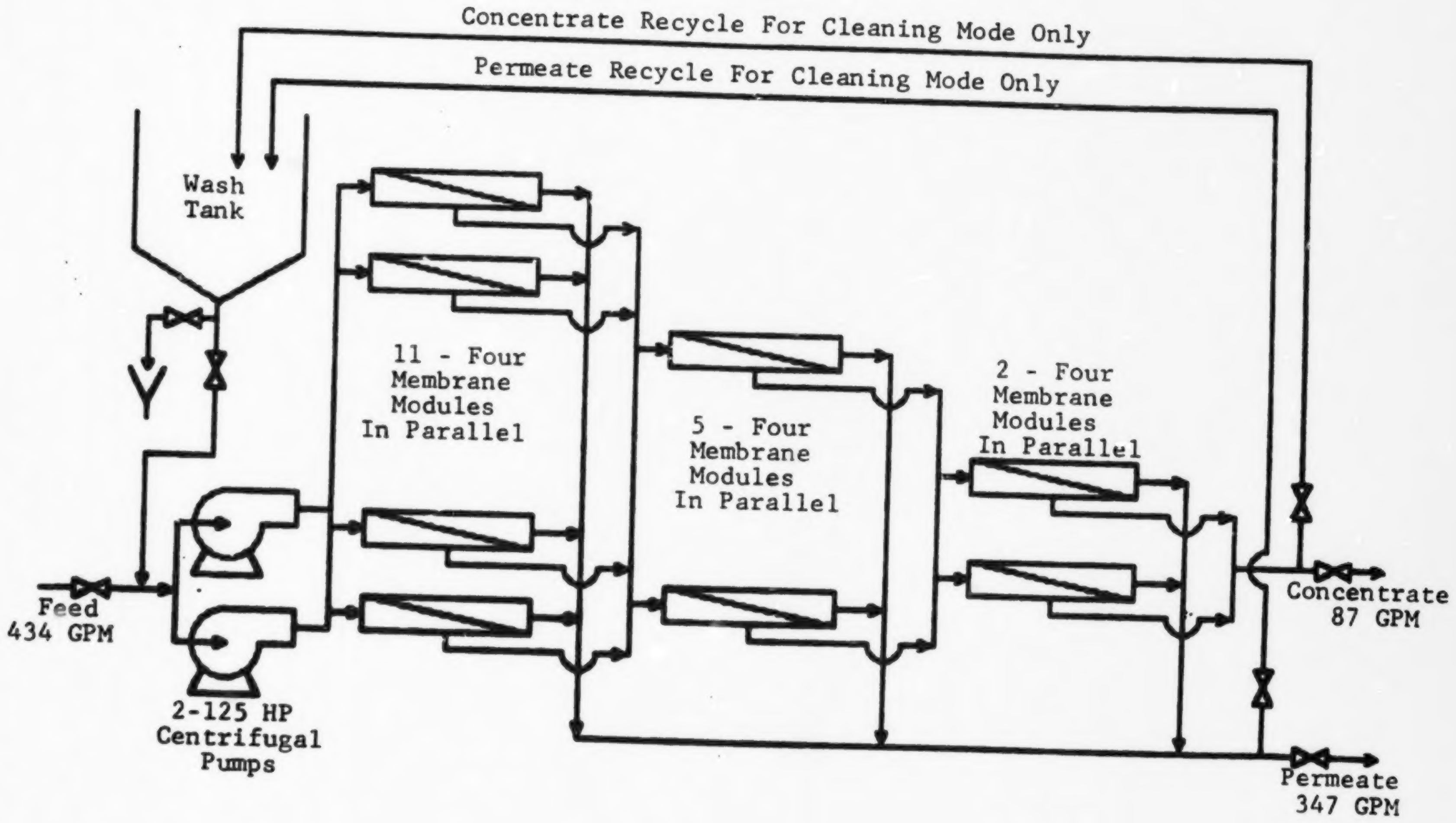
Number of membranes required = 72

18 four membrane modules will be required.

G.10.2 Equipment List and Purchase Costs

Purchase cost of one 72 membrane system (engineering and assembly work completed)	$480,298
Purchase cost of equipment	$480,298

FIGURE G-6 SPIRAL REVERSE OSMOSIS SYSTEM FOR DEMULSIFIED PRODUCED WATER.


Concentrate Recycle For Cleaning Mode Only
Permeate Recycle For Cleaning Mode Only
Wash Tank
11 - Four Membrane Modules In Parallel
5 - Four Membrane Modules In Parallel
2 - Four Membrane Modules In Parallel
Feed 434 GPM
2-125 HP Centrifugal Pumps
Concentrate 87 GPM
Permeate 347 GPM

G.10.3 Total Cost of Installed Equipment

Purchased cost of equipment		$ 480,298	
Shipping charges		24,015	
Installation of equipment		28,818	
Foundations		24,015	
Instruments		28,818	
Piping		24,015	
Insulation		33,621	
Electrical		38,424	
Buildings & structures		48,030	
Serivces		38,424	
Misc.		24,015	
Taxes	Federal	$79,249	
TOTAL DIRECT COSTS			$871,741
Construction and field expenses		43,587	
Engineering expenses		52,304	
Contractors fees		48,382	
Contingencies		96,763	
TOTAL INDIRECT COSTS			$ 241,036
TOTAL COSTS			$1,112,777

G.10.4 Operating and Maintenance Costs

Materials	
Cleaning Chemicals (wash every 2 days)	$25,550/year
Replacement Membranes (every 3 years)	$43,848/year
Labour (shared with other parts of plant)	
(1/2 shift/day at $135/day)	$16,875/year
Energy	
Electrical Power	$22,050/year
(250 HP x 0.7457 x 24 x 365 x 1.35 cents/KW)	
Miscellaneous	$4,000/year
Contingency	$11,200/year
Total Operating Costs	$123,523/year